

L Number	Hits	Search Text	DB	Time stamp
1	9	5073352.pn. 5525311.pn. 5600052.pn. 5600053.pn.	USPAT; US-PGPUB; EPO; JPO; DERWENT USPAT	2003/02/18 11:54
4	277	fast same fluidized same bed same reactor	USPAT	2003/02/18 11:56
8	21	("3928483" "4025575" "4252479" "4328384" "4447669" "4496786" "4499314" "4499327" "4547616" "4677242" "4677243" "4843183" "4849091" "4861938" "4873390" "4929780" "4973792" "5095163" "5126308" "5177279" "5191141").PN.	USPAT	2003/02/18 12:06
9	11	("3563911" "3753893" "3893812" "3919115" "3953175" "4197189" "4211630" "4234411" "4340566" "4371453" "4477335").PN.	USPAT	2003/02/18 12:00
10	11	("3563911" "3753893" "3893812" "3919115" "3953175" "4197189" "4211630" "4234411" "4340566" "4371453" "4477335").PN.	USPAT	2003/02/18 12:00
11	11	("3563911" "3753893" "3893812" "3919115" "3953175" "4197189" "4211630" "4234411" "4340566" "4371453" "4477335").PN.	USPAT	2003/02/18 12:00
14	1719	UOP adj inc	USPAT	2003/02/18 12:08
15	1184	(UOP adj inc) and catalyst	USPAT	2003/02/18 12:08

The Reference -
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United States Patent [19]

Cabrera et al.

[11] Patent Number: 4,849,091

[45] Date of Patent: Jul. 18, 1989

[54] PARTIAL CO COMBUSTION WITH STAGED
REGENERATION OF CATALYST

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[21] Appl. No.: 182,812

[22] Filed: Apr. 18, 1988

3,953,175 4/1976 Pulak 208/164
4,197,189 4/1980 Thompson et al. 208/164
4,211,630 7/1980 Gross et al. 208/164
4,234,411 11/1980 Thompson 208/164
4,340,566 7/1982 Thompson et al. 422/144 X
4,371,453 2/1983 Vickers 208/164
4,477,335 10/1984 Benslay 208/113

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[57] ABSTRACT

FCC catalyst is regenerated in a process providing increased coke-burning capacity without additional heat evolution or air consumption. The process uses a two-stage regeneration arrangement providing initial coke combustion in a low catalyst density-high efficiency contact zone followed by substantial separation of catalyst and regeneration gas and complete regeneration of catalyst particles in a dense bed regeneration zone. Catalyst and gas flow cocurrent prior to this separation but flow countercurrent after the separation. An effective control scheme for regulating oxygen addition to the final zone is also disclosed. This process is applicable to FCC operations for conventional and residual feedstocks.

11 Claims, 2 Drawing Sheets

Related U.S. Application Data

[63] Continuation-in-part of Ser. No. 908,531, Sep. 17, 1986,
abandoned.

[51] Int. Cl.⁴ C10G 13/18

[52] U.S. Cl. 208/113; 208/DIG. 1;
208/159; 208/153; 422/140

[58] Field of Search 208/113, 111, 153, 157,
208/148, 159, 163, 164, DIG. 1, 157; 422/140,
144

[56] References Cited

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- | | | | | |
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| 3,563,911 | 2/1971 | Pfeiffer et al. | | 422/144 X |
| 3,753,893 | 8/1973 | Wooyang Lee et al. | | 208/164 |
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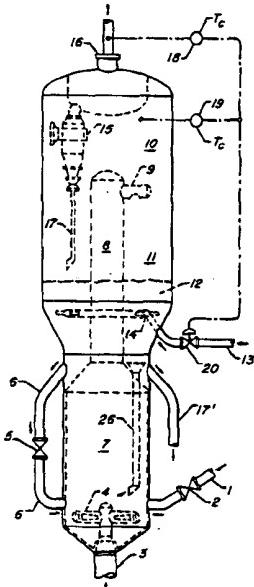


Figure 1

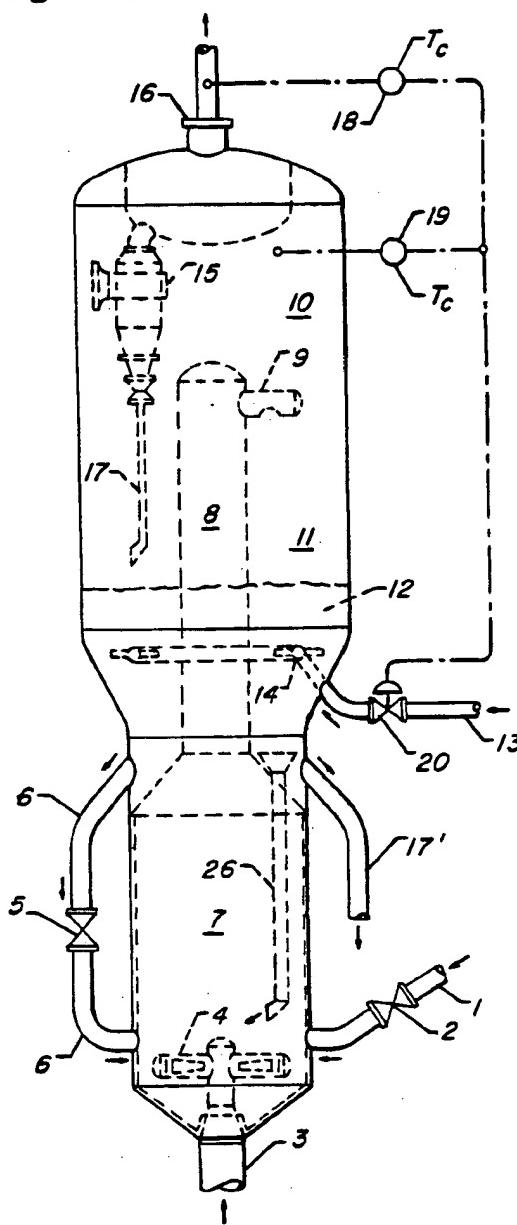


Figure 2

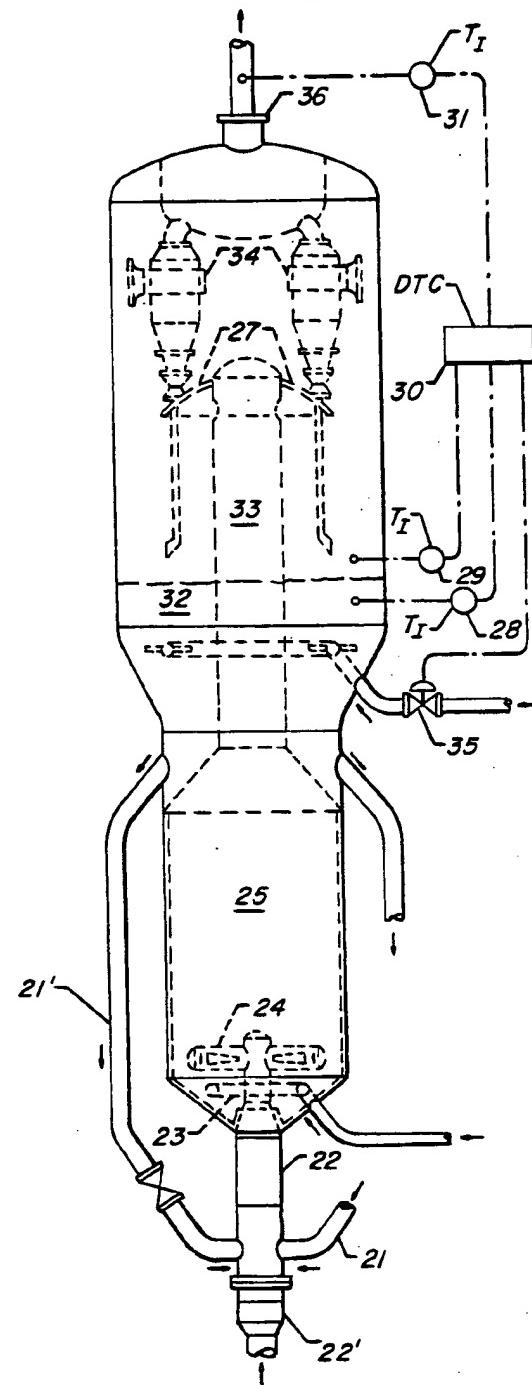
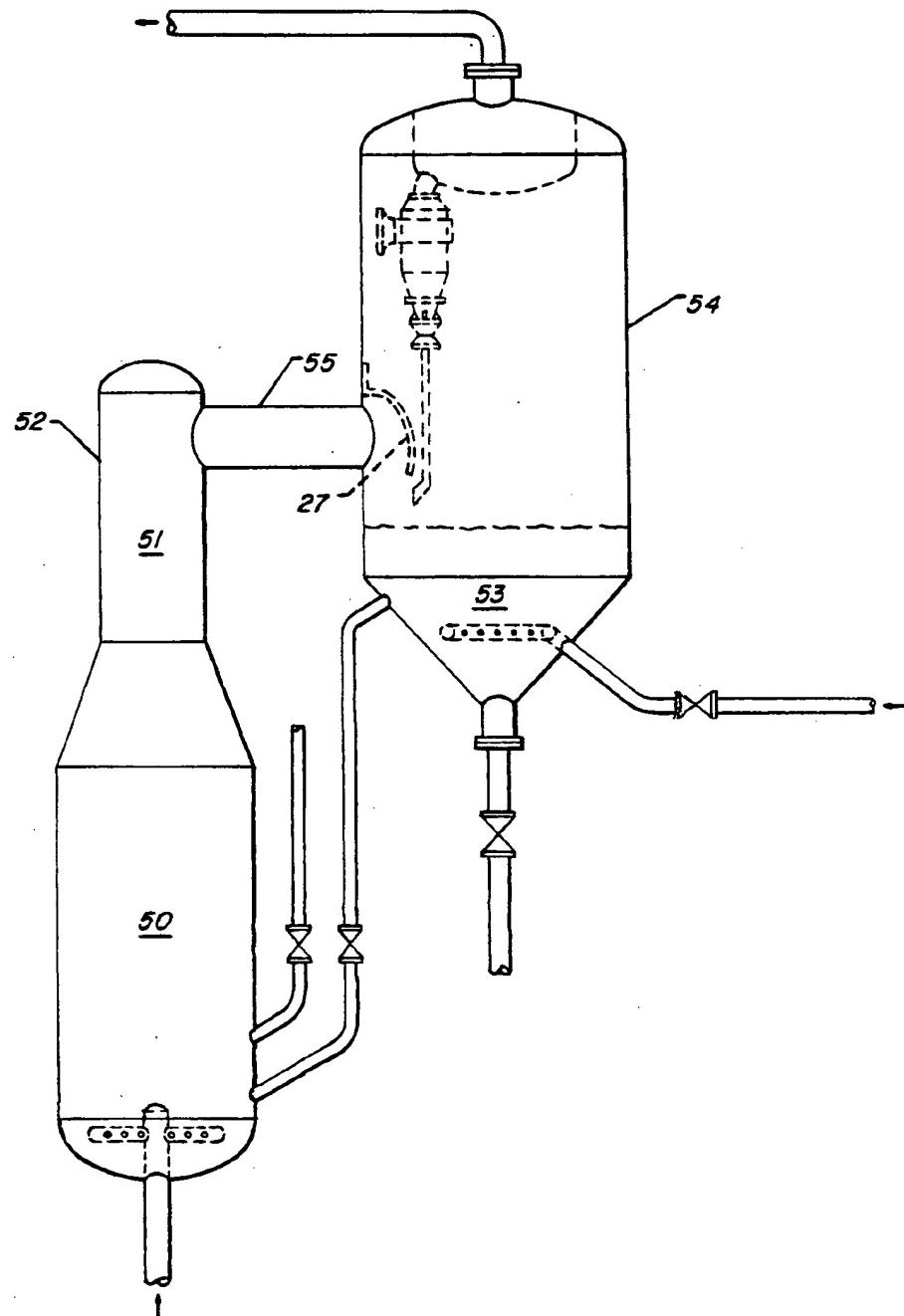


Figure 3



PARTIAL CO COMBUSTION WITH STAGED
REGENERATION OF CATALYST

CROSS-REFERENCE TO RELATED
APPLICATION

This application is a continuation-in-part of our prior copending application Serial No. 908,531 filed Sept. 17, 1986 abandoned.

BACKGROUND OF THE INVENTION

The invention relates to a method of regenerating used hydrocarbon conversion catalyst by the combustion of coke on the catalyst in a fluidized combustion zone. This invention specifically relates to a process for the conversion of heavy hydrocarbons into lighter hydrocarbons with a fluidized stream of catalyst particles and regeneration of the catalyst particles to remove coke which acts to deactivate the catalyst.

Fluidized catalytic cracking is a hydrocarbon conversion process accomplished by contacting hydrocarbons in a fluidized reaction zone with a catalyst composed of finely divided particulate material. The reaction in catalytic cracking, as opposed to hydrocracking, is carried out in the absence of added hydrogen or the consumption of hydrogen. As the cracking reaction proceeds substantial amounts of highly carbonaceous material referred to as coke is deposited on the catalyst. A high temperature regeneration within a regeneration zone operation burns coke from the catalyst. Coke-containing catalyst, referred to herein as spent catalyst, is continually removed from the reaction zone and replaced by essentially coke free catalyst from the regeneration zone. Fluidization of the catalyst particles by various gaseous streams allows the transport of catalyst between the reaction zone and regeneration zone. Methods for cracking hydrocarbons in a fluidized stream of catalyst, transporting catalyst between reaction and regeneration zones, and combusting coke in the regenerator are well known by those skilled in the art of fluidized catalytic cracking (FCC) processes. To this end the art is replete with vessel configurations for contacting catalyst particles with feed and regeneration gas respectively.

A common objective of these configurations is maximizing product yield from the reactor while minimizing operating and equipment costs. Optimization of feedstock conversion ordinarily requires essentially complete removal of coke from the catalyst. This essentially complete removal of coke from catalyst is often referred to as complete regeneration. Complete regeneration produces a catalyst having less than 0.1 and preferably less than 0.05 weight percent coke. In order to obtain complete regeneration, oxygen in excess of the stoichiometric amount necessary for the combustion of coke to carbon oxides is charged to the regenerator. Excess oxygen in the regeneration zone will also react with carbon monoxide produced by the combustion of coke thereby yielding a further evolution of heat.

When CO combustion occurs in a relatively catalyst free zone of the regenerator, such as the region above a dense fluidized bed, the resulting high temperatures may lead to severe equipment damage. Such situations may be avoided if the CO combustion takes place in the presence of catalyst particles which act as a heat sink. Therefore, regenerators are generally designed to avoid the combination of free oxygen and carbon monoxide in regions that are relatively free of catalyst. Despite this

the heat evolved from unintended CO combustion may raise the temperature of the catalyst to the point of causing thermal deactivation of the catalyst or may affect the process by limiting the amount of catalyst that can contact the feedstock. The problems of controlling catalyst and regenerator temperatures are exacerbated by the application of FCC processes to crack heavy feedstocks. With the increased coke producing tendencies of these heavy or residual feeds a complete regeneration of catalyst becomes more difficult due to the excessive heat evolution associated with coke and CO combustion.

Aside from excessive heat evolution the complete oxidation of CO also increases overall oxygen demands for the regeneration process. In many cases the high oxygen requirements for complete regeneration may exceed the limited capacity of the regenerator air blower or other equipment in the regenerator and flue gas section. Therefore, it is also desirable to reduce CO combustion so that a greater quantity of the available air supply is used to oxidize coke from the catalyst.

INFORMATION DISCLOSURE

One way to minimize CO combustion and yet obtain fully regenerated catalyst is by performing the regeneration in stages. Staged regeneration systems are well known in the regeneration of FCC catalyst. Luckenbach, U.S. Pat. No. 3,958,953, describes a staged flow system having concentric catalyst beds separated by baffles which open into a common space for collecting spent regeneration gas and separating catalyst particles. Myers et al. in U.S. Pat. No. 4,299,687 teach the use of a staged regenerator system having superimposed catalyst beds wherein spent catalyst particles first enter an upper dense fluidized bed of catalyst and are contacted with regeneration gas from the lower catalyst bed and fresh regeneration gas. After partial regeneration in the first regeneration zone, catalyst particles are transferred by gravity flow into a lower catalyst bed to which is charged a stream of fresh regeneration gas. The Myers invention is directed to the processing of residual feeds and uses the two stage regeneration process to limit CO combustion thereby reducing overall heat output within the regenerator.

The use of relatively dilute phase regeneration zones to effect complete catalyst regeneration is shown by Stine et al. in U.S. Pat. Nos. 3,844,973 and 3,923,686. Stine et al. seeks primarily to effect complete CO combustion for air pollution, thermal efficiency, and equipment minimization reasons by using increased gas velocities to transport catalyst through dense bed and relatively dilute phase regeneration zones. A two stage system which combines a relatively dilute phase transport zone with a dense bed zone for regenerating catalyst used in cracking residual feeds is shown by Dean et al. in U.S. Pat. No. 4,336,160. In Dean a first dense bed is used to initiate coke combustion in a lower portion of a regeneration section which is followed by an upper dilute phase regeneration section operating at high severity to complete regeneration and combustion of carbon monoxide.

In the various methods of practicing staged regeneration it is necessary to supply fresh regeneration gas to each regeneration zone. Control of the regeneration gas to each zone allows the degree of coke combustion and CO oxidation to be determined for each stage. Therefore, it is desirable to minimize the interdependence of

gas flow rates upon different stages of regeneration. Multistage regeneration systems in which spent or partially spent regeneration gas from one regeneration stage enters another stage have flow rate and oxygen concentration constraints that interfere with the control of CO combustion. Although allowing the optimum in control, essential isolation of the regeneration gas streams results in higher equipment costs due to the need for extra separation devices and piping.

Partial isolation of a regeneration gas stream entering a final regeneration zone is taught by Benslay in U.S. Pat. No. 4,477,335. This reference also shows the use of a single catalyst separation section for all regeneration stages located in the final regeneration zone. However, this method and apparatus uses multiple dense beds with an unusual downflow riser to transfer catalyst to a final stage located at the lowermost portion of the regenerator.

U.S. Pat. No. 3,563,911 issued to R. W. Pfeiffer et al. illustrates a multistage FCC catalyst regeneration zone. The spent catalyst is fed into a dense bed of fluidized catalyst. Each stage appears to be operated at similar conditions. Catalyst of various degrees of regeneration collected in the common vapor volume above the horizontally contiguous regeneration stages is returned to the first regeneration stage. The oxygen supply to the regeneration zone is controlled to provide a small amount of afterburning.

U.S. Pat. No. 3,953,175 issued to R. P. Pulak is believed pertinent for its showing that catalyst can be regenerated in a vertical chamber having a smaller diameter section which acts as a riser to transfer regenerated catalyst. This process appears to be directed to the complete combustion of CO to CO₂.

U.S. Pat. No. 4,197,189 issued to G. J. Thompson et al. illustrates a process for the regeneration of spent FCC catalyst wherein the catalyst flows upward through a cylindrical combustion zone and is then discharged in a manner providing for gas-catalyst separation. All of the regeneration oxygen appears to enter this combustion zone, which therefore is believed to be properly described as a single stage of combustion despite dual air feed points. Regenerated hot catalyst is collected and partially recycled to the bottom of the combustion zone.

In contrast to the prior art the present invention uses fast fluidized flow conditions, hereinafter described, in a first stage of regeneration to provide a highly efficient contacting of catalyst and oxygen which will minimize the presence of free oxygen in the spent regeneration gas from the first regeneration stage thereby facilitating control of CO combustion in subsequent stages of the regeneration process. The catalyst and combustion gas flow upward concurrently prior to a gas-solids separation. Subsequent catalyst flow is countercurrent to rising gas emanating from a dense bed of catalyst.

OBJECTS AND EMBODIMENTS

It is an object of this invention to provide a process for regenerating catalyst with minimum oxygen input requirements and minimum heat output. It is a further object of applicants' invention to maximize the flexibility of regeneration operations within a two-stage regeneration system with minimum equipment demands. These objectives are met using a dual combustion stage process featuring intermediate separation of the catalyst and regeneration gases. The oxygen availability and

other operational characteristics differ between the stages.

Accordingly, a broad embodiment of the present invention is a process for the regeneration of spent hydrocarbon conversion catalyst containing coke which has been removed from a fluidized catalytic reaction zone, which process comprises the steps of passing to a lower locus of a combustor zone of a regeneration zone spent catalyst from said reaction zone, a stream consisting of regenerated catalyst from a hereinafter described dense bed regeneration zone, and a first oxygen containing regeneration gas stream in an amount sufficient to maintain fast fluidized conditions within said combustor zone and to oxidize coke along with coke combustion by-products; oxidizing coke and coke combustion by-products in said combustor zone while transporting said spent and regenerated catalyst upward in cocurrent flow with rising regeneration gas; passing at least a portion of said spent and regenerated catalyst and regeneration gas upward in cocurrent flow from the combustor zone into a riser regeneration zone located above said combustion zone and having a lower catalyst density than said combustor zone, and therein oxidizing coke and coke combustion by-products to produce partially regenerated catalyst and a spent first regeneration gas having a mole ratio of CO₂/CO of from about 0.7 to about 2.0; discharging partially regenerated and regenerated catalyst, and said spent first regeneration gas from an upper locus of said riser regeneration zone into a catalyst disengagement zone through an outlet means that effects at least a partial separation of catalyst and regeneration gas and thereby causing an initial separation of catalyst from said spent first regeneration gas; allowing at least 70 wt. % of the partially regenerated and regenerated catalyst discharged through said outlet means to settle into a dense fluidized bed of a dense bed regeneration zone located below said disengagement zone while flowing countercurrent to a rising second regeneration gas which contains at least 0.5 mole percent oxygen, and introducing into the dense fluidized bed a second oxygen containing regeneration gas stream in a quantity at least sufficient to produce regenerated catalyst having less than 0.1 wt. % coke and to oxidize essentially all of the carbon monoxide produced within said dense bed regeneration zone to carbon dioxide; combining said spent first regeneration gas with the second regeneration gas, which emanates from said dense bed regeneration zone in said disengagement zone, and producing a combined spent regeneration gas having a CO₂/CO mole ratio in the range of about 1 to 5; withdrawing said combined spent regeneration gas from the disengagement zone; and withdrawing regenerated catalyst from said dense bed having an average temperature in the range of 620 to 787 degrees Celsius and returning at least a first portion of said regenerated catalyst to said reaction zone and at least a second portion of regenerated catalyst to the lower locus of the combustor zone as set out above.

In an alternative embodiment the invention is a process for the regeneration of particulate hydrocarbon cracking catalyst withdrawn from a fluidized catalytic cracking reaction comprising the steps of passing a stream comprising spent catalyst from said reaction zone to a lower locus of a combustion zone together with regenerated catalyst from a hereinafter described dense bed regeneration zone and an oxygen containing regeneration gas stream in sufficient quantity to obtain fast fluidized conditions within said combustion zone

and to oxidize coke and coke combustion by-products including carbon monoxide; oxidizing coke and coke combustion by-products in said combustion zone while transporting catalyst upward cocurrent with regeneration gas; passing catalyst and regeneration gas mixture to a riser regeneration zone located above said combustion zone and operated at lower catalyst density than said combustion zone and therein oxidizing coke and coke combustion by-products to produce partially regenerated catalyst and a spent (oxygen-depleted) first regeneration gas having a mole ratio of CO₂/CO of from about 0.7 to 2.0; discharging said partially regenerated catalyst, recirculated regenerated catalyst and spent first regeneration gas from an upper locus of said riser regeneration zone into a catalyst disengagement zone through an outlet means that effects at least a partial separation of catalyst and regeneration gas and thereby causing an initial separation of partially regenerated catalyst from said first regeneration gas; collecting at least a major portion of said partially regenerated and recirculated catalyst discharged into the catalyst disengagement zone through said outlet in a dense bed regeneration zone located below said disengagement zone, introducing into the dense bed regeneration zone a second oxygen containing regeneration gas stream at a rate which is adjusted in response to a hereinafter defined output signal; combining said spent first regeneration gas with a second regeneration gas which rises upward from said dense bed regeneration zone in said disengagement zone countercurrent to descending catalyst to obtain a combined regeneration gas having a CO₂/CO mole ratio of from 1 to about 5 and which is greater than the mole ratio of CO₂/CO of said spent first regeneration gas, separating entrained catalyst from the combined spent regeneration gas, removing the combined regeneration gas from said disengagement zone and returning the entrained catalyst to said dense bed regeneration zone; measuring the temperature of the combined regeneration gas, comparing said temperature to a predetermined or variable set point, and generating from this comparison an output signal which determines the flow rate of oxygen containing regeneration gas stream to the dense bed regeneration zone such that said set point is not exceeded; and withdrawing regenerated catalyst from said dense bed regeneration zone having an average temperature in the range of 620-787 degrees Celsius (1150 to 1450 degrees Fahrenheit) and returning said regenerated catalyst particles to said reaction zone.

BRIEF DESCRIPTION OF THE DRAWINGS

FIG. 1 is a cross sectional view of a suitable regenerator for performing the subject process.

FIGS. 2 and 3 illustrate alternative regenerator configurations suitable for employing the invention.

DETAILED DESCRIPTION OF THE INVENTION

This invention is directed to an arrangement for removing carbon deposits referred to as coke from the surface and pores of catalyst used in a hydrocarbon conversion process. A preferred embodiment of the invention is a two-stage FCC regeneration process. The first regeneration stage utilizes fast fluidization conditions in a high efficiency regeneration stage for partially regenerating spent hydrocarbon cracking catalyst while yielding a CO rich spent regeneration gas. A second

regeneration stage uses a dense bed to complete regeneration of the catalyst.

The completely regenerated catalyst of this process is obtained with a significantly reduced heat output and with a lower oxygen requirement per pound of coke burned than required to obtain similar catalyst from regenerators not incorporating the features of the present invention. Reducing the heat output allows the burning of additional coke within the regenerator without a subsequent increase in regenerator temperatures. Furthermore lower oxygen requirements per pound of coke burned permits burning additional coke for a given oxygen supply.

Alternately the invention may be employed to lower the temperature of the regenerated catalyst thereby allowing operational changes in the reaction zone. It has been recognized that high regeneration temperatures of past regeneration schemes did not provide catalyst at optimum condition for maximizing the yield of desired products from the reaction zone. As hereinafter demonstrated increased circulation of fully regenerated catalyst as a result of lower regenerated catalyst temperature offers definite advantages in maximizing product yield. Accordingly, the ability to obtain fully regenerated catalyst in an optimum temperature range is an important benefit of this process. By means of the present invention the temperature of the regenerated catalyst is easily controlled while obtaining complete regeneration of the catalyst in a highly efficient manner.

Referring now to the accompanying drawings the regeneration process of the present invention will be described in more detail. In FIG. 1 spent coke-containing catalyst from a reaction zone enters the lower portion of combustion zone 7 through conduit 1 containing control valve 2. The catalyst from the reactor usually contains carbon in an amount of from 0.2 to 2 wt. %, which is present in the form of coke. Although coke is primarily composed of carbon it may contain from 3 to 15 wt. % hydrogen as well as sulfur and other materials. An oxygen containing regeneration gas, typically air, enters a lower portion of the combustion zone through conduit 3 and is distributed across combustor zone 7 by distributor 4. As the regeneration gas enters the combustion zone it contacts spent catalyst.

In order to accelerate combustion of the coke, hot regenerated catalyst from dense bed regeneration zone 12 may be recirculated into the combustor zone via conduit 6 which contains control valve 5. Recirculation of regenerated catalyst, by mixing hot catalyst from dense bed regeneration zone 12 with relatively cold spent catalyst entering the combustor zone, raises the overall temperature of the catalyst and gas mixture in combustion zone 7. Aside from external standpipe 6, several other methods of effecting catalyst recirculation are well known. For instance, catalyst may be transferred internally by internal standpipe 26.

The catalyst and gas mixture then enter riser regeneration zone 8, which is operated at a higher gas velocity due to its reduced cross section. The primary function of the first regeneration stage comprising combustion zone 7 and riser regeneration zone 8 is to maximize coke combustion to carbon monoxide while limiting the combustion of CO to CO₂ thereby minimizing oxygen consumption per unit of coke burned. Typically the amount of coke removed in this first regeneration stage comprises from about 50% to 90% of the coke on entering spent catalyst. The minimization of CO combustion to CO₂ results in an increase in capacity for burning addi-

tional coke deposits and at the same time minimizes heat evolution and air requirements during regeneration. The additional coke burning capacity increases the flexibility of the reaction zone in processing feeds having increased coke producing tendencies.

The mixture of catalyst particles and regeneration gas, which is spent due to oxygen consumption, is discharged from an upper portion of riser zone 8. Both totally regenerated recirculated catalyst and partially regenerated catalyst exit the top of riser zone 8. Discharge is effected through disengaging device 9 which separates a majority of the partially regenerated catalyst from the spent regeneration gas. Initial separation of catalyst upon exiting riser zone 8 minimizes the catalyst loading on cyclone separators or other downstream devices used for the essentially complete removal of catalyst particles from the spent regeneration gas thereby reducing overall equipment costs. This initial separation also makes possible the operation of the upper section of the vessel at higher gas velocities than would be possible in the case of a conventional fluidized bed. There are various flow devices known to those skilled in the art, that will perform the preliminary catalyst and gas separation any of which would be suitable for use in this invention as disengaging device 9. Following the initial catalyst and gas separation, regeneration gas and a minor portion of the catalyst still entrained therein will rise into the upper portion of a disengaging space 10. A major portion, at least 70 and preferably 80 wt. %, of the catalyst now disengaged from the regeneration gas falls through the disengaging space 11 countercurrent to regeneration gases.

Downward moving disengaged catalyst collects in dense bed regeneration zone 12. Catalyst densities in this zone are typically kept within a range of from about 480-800 Kg per cubic meter (30 to 50 pounds per cubic foot). A second quantity of oxygen-containing regeneration gas typically air, enters this dense bed regeneration zone through conduit 13 and distribution device 14. Approximately 10 to 50 percent of the total oxygen requirements within the process for the essentially complete regeneration of the catalyst enters dense bed regeneration zone 12. Although the total quantity of regeneration gas charged to dense bed regeneration zone 12 represents half or less of the total regeneration gas to the process it will be adequate to provide this amount of regeneration gas since catalyst entering the dense bed regeneration zone contains less than half of its original coke deposits. The oxygen requirements for complete regeneration in the dense zone are commensurately reduced. Therefore, catalyst in the dense bed regeneration zone is completely regenerable to a carbon content of less than 0.1 wt. %.

The flow of catalyst and gas is cocurrent in the riser regeneration zone. After discharge from the riser, the catalyst flows countercurrent to rising gases. The highest oxygen concentration is therefore present at different ends of the two combustion zones. In the riser regeneration zone, oxygen concentration is highest at the point of greatest carbon content on the catalyst. In the countercurrent disengaging area and in the dense bed regeneration zone, the highest oxygen level exists in unison with the most highly regenerated catalyst.

Regeneration gas containing at least 0.5 mole percent oxygen rises from the bed and passes upward through disengagement space 11. The downward discharge of disengaged catalyst from separation device 9 increases the concentration of catalyst in the dilute phase above

the dense bed thereby providing an additional region in which coke and CO oxidation may take place as the falling catalyst particles contact the oxygen containing regeneration gas. As the regeneration gas from the dense bed regeneration zone 12 continues to rise it is combined with regeneration gas from the first combustion zone in upper disengaging space 10. These two streams upon first contact have a combined CO_2/CO mol ratio of between 1 to 5 and of course greater than the same ratio in the regeneration gas exiting the riser. Oxygen present in the regeneration gas from the dense bed regeneration zone may cause oxidation of carbon monoxide in the upper disengaging space. Although the overall heat output and final regeneration gas temperature will increase, this burning of coke and CO in the dilute phase above the dense bed and in the disengaging zone will not cause a major increase in catalyst temperatures due to the amount of catalyst present. Therefore, the invention contemplates a variation in temperatures for the regeneration gas in the various zones. For example, when carbon monoxide is combusted in the upper portion of the disengagement zone the differential between the temperature of the riser regenerator gas entering the disengagement zone and the combined regeneration gas can reach from 27.8-111 Celsius degrees (50-200 Fahrenheit degrees).

The combined regeneration gas stream and entrained particles of catalyst enter one or more separation means, such as cyclone separator 15, which separates catalyst fines from the gas stream. Regeneration gas, relatively free of catalyst is withdrawn from the regenerator through outlet 16 while recovered catalyst is returned to the dense bed zone through dip leg 17 or other comparable means. From about 10 to 30 wt. % of the catalyst discharged from the riser regeneration zone is present in the gases above the exit from the riser regeneration zone and enter the cyclone separator. Catalyst from the dense bed regeneration zone is transferred through line 17' back to the reactor where it again contacts feed as the process continues.

For the purposes of this invention the term high efficiency regeneration refers to the use of fast fluidized flow conditions within an FCC regeneration section. Fast fluidization defines a condition of fluidized solid particles lying between the turbulent bed of particles and complete particle transport mode. A fast fluidized condition is characterized by a fluidizing gas velocity higher than that of a dense phase turbulent bed, resulting in a lower catalyst density and vigorous solid/gas contacting. In a fast fluidized zone, there is a net transport of catalyst caused by the upward flow of fluidizing gas. The catalyst density in the fast fluidized condition is much more sensitive to particle loading than in the complete particle transport mode. Therefore, it is possible to adjust catalyst residence time to achieve the desired combustion at the highly effective gas-solid, mixing conditions. From the fast fluidized mode, further increases in fluidized gas velocity will raise the rate of upward particle transport, and will sharply reduce the average catalyst density until, at sufficient gas velocity, the particles are moving principally in the complete catalyst transport mode. Thus, there is a continuum in the progression from a fluidized particle bed through fast fluidization and to the pure transport mode.

In this invention the combustion zone 7 will have a catalyst density of from 48 to 400 kg per cubic meter (3 to 25 pounds per cubic foot) and superficial gas velocities from about 0.91 to 3.05 meters per second (3 to 10

feet per second). Riser regeneration zone 8 is operated at a higher gas velocity which will usually exceed 3.05 meters per second (10 feet per second) and, therefore, will have a flow regime at the upper limit of fast fluidization conditions or the lower limit of the essentially pure transport mode. As a result the riser regeneration zone will have lower catalyst densities of from 16 to 128 kg per cubic meter (1 to 8 pounds per cubic foot), higher transport rates and reduced backmixing as compared to the combustor zone. Nevertheless the riser and combustion zones provide regions of lower catalyst density and vigorous mixing which are believed to be the most efficient for coke combustion and characterize a high efficiency regeneration. Therefore, the use of a first high efficiency contact assures removal of a major portion of coke from the catalyst in the initial regeneration stages. The addition of regeneration gas at conditions to promote high efficiency regeneration is sufficient to remove more than 50 percent and preferably between 65 and 90 wt. percent of the coke from the entering spent catalyst in the combustor and riser zones.

When oxidizing the carbon components of coke from the catalyst there are three principal reactions that can take place:



Thus in oxidizing the carbon either carbon monoxide or carbon dioxide is formed. The reaction equations show that maximizing the production of carbon monoxide while reducing the production of carbon dioxide minimizes the release of heat during regeneration. This minimization of heat release is substantial since less than one third as much heat is released in forming carbon monoxide than is released in forming carbon dioxide. Of course, it is ordinarily necessary to provide more than the stoichiometric requirement of oxygen for conversion of the carbon products in order to completely regenerate the catalyst. Thus complete regeneration requires sufficient excess oxygen which also reacts with carbon monoxide to yield carbon dioxide and additional heat. By restricting the oxygen concentration in a regeneration zone it is possible to suppress carbon monoxide conversion, but unfortunately coke removal is reduced at the same time.

In the subject invention the use of fast fluidization conditions provides optimum conditions for carbon removal and the complete utilization of oxygen. Therefore, it is possible to obtain a high percentage reduction of coke on the catalyst while still keeping the ratio of $CO_2:CO$ in first combustion zone in the range of 0.7 to 2.0. Maintaining a low ratio of $CO_2:CO$ while oxidizing a major portion of the coke lowers the overall temperature of the regeneration process and yields an oxygen deficient spent regeneration gas.

An additional benefit of reducing carbon monoxide combustion to carbon dioxide is a corresponding reduction in the total oxygen requirements for catalyst regeneration. Although the hereinafter described dense bed regeneration zone will also present a significant demand for oxygen, the overall oxygen requirements of the invention are significantly reduced. This reduction in oxygen requirements can be used to increase the coke burning capacity of a regenerator zone. Providing additional coke burning capacity without extra air is partic-

ularly useful where the coke burning capacity of the regenerator is limited by the size of the air blower.

Moreover the process of this invention is arranged in a manner that facilitates control of coke combustion and regulation of regeneration gas. First, each stream of regeneration gas only contacts the catalyst in one regeneration zone. Hence, regeneration gas from one zone does not interfere with the operations taking place in another zone. In stacked regeneration zones of the past the passage of regeneration gas from one zone to another zone interferes with control of coke combustion or regeneration temperature. For example, where a regeneration gas containing oxygen from a final combustion stage is circulated through a first combustion stage it is not possible to control the afterburning of CO to CO_2 by the quantity of free oxygen supplied to the final combustion stage.

By performing the first regeneration step in a high efficiency contact zone a greater utilization of available oxygen can be obtained in the first stage, enhancing the ability to control regeneration in the dense bed regeneration zone. As an aside, the mixing of partially spent regeneration gas with fresh regeneration gas in the subsequent dense bed regeneration zone, as practiced in the art, increases the superficial velocity through subsequent dense bed regeneration zones. Increased superficial velocity results in increased catalyst entrainment. This in turn requires separation devices of increased capacity. Of course, it is possible to provide individual separation means for each regeneration zone, thereby avoiding the disadvantage of passing spent regeneration gas through subsequent zones. However, such a step requires additional separation equipment. This invention combines the advantages of a single catalyst separation and flue gas section with lower catalyst entrainment while operating at higher overall gas rates.

Thus an FCC reaction zone associated with this invention can be used to process a conventional FCC feedstock or higher boiling hydrocarbon feeds. The most common of such conventional feedstocks is a "vacuum gas oil" (VGO), which is typically a hydrocarbon material having a boiling range of from 343 to 552 degrees Celsius (650 to 1025 degrees Fahrenheit) prepared by vacuum fractionation of atmospheric residue. Such a fraction is generally low in coke precursors and heavy metal contamination which can serve to contaminate catalyst.

Heavy hydrocarbon feedstocks to which this invention may be applied include heavy bottoms from crude oil, heavy bitumen crude oil, shale oil, tar sand extract, deasphalted residue, products from coal liquefaction, atmospheric and vacuum reduced crudes. Heavy feedstocks for this invention also include mixtures of the above hydrocarbons. However, the foregoing list is not intended to exclude the application of this process to other suitable feeds.

The heavy hydrocarbon fractions are also characterized by the presence of significant metal contamination. These metals accumulate on the catalyst and poison the catalyst by blocking reaction sites and promote over-cracking thereby interfering with the reaction process. Therefore, the use of passivation or other metals management procedures within or before the reaction zone are anticipated when processing heavy feedstocks by this invention.

Therefore, one advantage of the process is that it allows the processing of heavier feedstocks in an exist-

ing unit with only a minor revamp or reduces the cost of a new unit designed to process heavier feeds. This benefit is a direct result of the increased coke burning capacity of the unit which can be attributed to the lower heat outputs and oxygen requirements for combustion per unit of coke. With regard to oxygen or air requirements, a typical high efficiency regenerator may require 14 kilograms of air per each kilogram of coke removed. By contrast, in the present invention complete regeneration may be obtained using as little as 11 kilograms of air per kilogram of coke.

It is also readily apparent that the present invention alternatively enables the processing of greater amounts of feed in a conventional reaction zone that employs the regeneration process of this invention than a process not employing the invention. However, the method of this invention can also improve the reactor performance when the quality or quantity of feed remains the same. This is a result of lower heat of combustion which leads to a lower catalyst temperature and greater catalyst circulation and with a commensurate increase in conversion.

Other embodiments of this invention are illustrated in FIG. 2 which depicts a somewhat modified regenerator configuration. The lower section of FIG. 2 illustrates a separate mixing zone for combining spent catalyst, regenerated catalyst and regeneration gas. In this configuration hot regenerated catalyst transported down extended standpipe 21' meets spent catalyst entering mixing riser 22 through conduit 21. Spent and regenerated catalyst are contacted with at least a portion of a first stream of oxygen containing regeneration gas from conduit 22' at a lower portion of mixing riser 22. The mixing riser has a reduced cross-sectional area in relation to the lower portion of the combustion zone to promote intimate mixing of the catalyst particles and gas stream. After mixing, the catalyst and gas mixture enter the lower portion of the first combustion zone 25 through distribution device 24 where it may be contacted with additional regeneration gas through inlet device 23. The operation of a mixing riser is more fully described by Thompson et al. in U.S. Pat. No. 4,340,566. FIG. 2 also depicts pipe arm separation device 27 as one possible alternative to separation device 9 of FIG. 1.

FIGS. 1 and 2 show a symmetrical configuration of the regeneration zones with the dense bed regeneration zone located above the combustor regeneration zone. However as demonstrated in FIG. 3, the riser 51 and combustor 50 zones may be contained in a separate vessel 52 and located adjacent to vessel 54 containing dense bed regeneration zone 53. Catalyst in this embodiment is transferred from the first regeneration stage to the dense bed regeneration zone by means of a horizontal transport riser 55. Thus the utilization of this invention is not limited to a symmetrical regenerator configuration but may be added to dense bed regeneration zones via the addition of a combustor, riser and transport conduit as taught by U.S. Pat. No. 3,953,175.

This invention also lends itself to a simple control method for regulating the addition of fresh regeneration gas to the dense bed regeneration zone. When operating a single stage dense bed regeneration zone it has been the practice to limit excess oxygen in the region above the dense bed to prevent so-called afterburning of carbon monoxide outside the bed of catalyst particles. Control of afterburning is also an objective of this invention due to the increased presence of carbon monox-

ide in the disengaging zone. It is possible to control the addition of regeneration gas to the dense bed zone by monitoring the spent regeneration gas temperature. In the control scheme of FIG. 1 the addition of regeneration gas to the dense bed regeneration zone is regulated by control valve 20 in response to a signal derived from either temperature controller 19, which senses the temperature in the upper portion of the disengagement zone, or from temperature controller 18 which senses the exiting regenerator gas temperature. The temperature controller may be set to keep the upper regenerator temperatures below the maximum equipment temperature.

Afterburning may also be employed beneficially as a means to monitor required oxygen addition rates or control regeneration gas addition to the dense bed zone. By measuring the temperature of the dense bed regeneration zone, the regeneration gas leaving the dense bed zone, the spent regeneration gas leaving the riser regeneration zone, or the two regeneration gas streams at the point of initial mixing, and comparing it to the temperature of the combined regeneration gas at a downstream location such as the upper portion of the disengagement space or spent regeneration gas outlet, the resulting differential temperature will indicate the occurrence of afterburning in the upper disengaging zone and the presence of oxygen in the disengagement zone. Assuming the fast fluidized zones are operated in an oxygen deficient mode, the presence of oxygen at this point will in turn insure that complete regeneration is occurring in the dense bed zone.

A suitable arrangement for this control is shown in FIG. 2 wherein a first signal representing either the dense bed temperature as measured by temperature indicator 28 or the dense bed regenerator gas temperature as measured by temperature indicator 29 is compared in differential temperature controller 30 to a second signal representing the temperature of the flue gas as measured by temperature indicator 31 in order to generate a differential temperature value. The differential temperature controller then generates, based on the differential temperature value, a third signal which is sent to control valve 35 to regulate addition of regeneration gas to the dense bed zone. Due to the high concentration of carbon monoxide in the disengaging space, the low heat capacity of the regeneration gas and the usually small proportion of regeneration gas added to the dense bed regeneration zone, this differential control means is highly responsive. Accordingly, the regeneration gas addition to the second zone may be adjusted to maintain a predetermined temperature differential thereby insuring the presence of excess oxygen in the second combustion zone and complete regeneration, but at the same time limiting the gas flow to avoid any excessive temperature differential or regeneration gas addition.

It is foreseeable that relatively cooler regeneration gas will enter the disengagement zone from the riser. Therefore, it is possible to base control on a negative temperature differential between the second and first signal of as much as 20 Celsius degrees. However, a minimum positive differential of 11 Celsius degrees is usually required to insure the presence of oxygen, with 17 Celsius degrees being preferred. Of course, the maximum temperature differential will be dictated by the temperature limitations of the catalyst and equipment. Thus for most FCC or reduced crude conversion operations the maximum temperature is about 787 degrees

Celsius (1450 degrees Fahrenheit) which will limit the differential temperature accordingly. However, the maximum positive temperature differential will usually not exceed 42 Celsius degrees where the regenerator is operated to minimize air requirements.

The following examples are presented to demonstrate the reduced heat release obtained by the process of this invention as well as a corresponding benefit to the reaction section. Nevertheless these examples represent only one possible method of practicing this invention 10 and are not meant to restrict the broad scope of the claims appended hereto. Furthermore, these examples incorporate engineering calculations and estimates, based on operating data which reliably reflect actual 15 operation.

EXAMPLE I

A high efficiency regenerator is operated without an additional dense bed regeneration zone to demonstrate the prior art process for regenerating coke containing catalyst from an FCC reactor hereinafter described in Example II. For purposes of illustration, reference will be made to the applicable elements of FIG. 2. Of course in this example there is no secondary addition of air to a dense bed regeneration zone.

The regenerator of this example, apart from the influence of catalyst loadings, operates at a pressure of 227 k pa gauge. In this operation 1,354,634 kg per hour of coke contaminated (spent) catalyst containing 0.85 wt. of coke and having a temperature of 525 degrees Celsius (977 degrees Fahrenheit) is transferred from an FCC reaction zone hereinafter described in Example II to mixing riser 22 through standpipe 21 where it is combined with 162,374 kg per hour of 163 degrees Celsius (325 degrees Fahrenheit) air and 1,354,634 kg per hour 35 of regenerated catalyst having a temperature of 739 degrees Celsius (1362 degrees Fahrenheit) and a coke content of less than 0.05 wt. %, taken from upper collection zone 32 through standpipe 21'. Thus a total of 11,514 kg/hr. of coke is charged to the regenerator. The 40 density of the catalyst and gas-catalyst mixture varies from 48-320 kg/m³ in the lower combustion zone to 16-80 kg/m³ in the combustion riser zone.

After passage through the combustion riser the mixture of catalyst and gas is initially separated at riser 45 outlet 27. Spent regeneration gas and entrained catalyst are further separated in cyclones 34. Spent regeneration gas at a temperature of about 743 degrees Celsius (1370 degrees Fahrenheit), essentially free of carbon monoxide, and containing 2 mole percent oxygen leaves the 50 regenerator through outlet 36. Regenerated catalyst particles from the riser outlet and cyclones collect in zone 32. Zone 32 contains a semifluidized bed of catalyst having a coke content of less than 0.05 percent and a temperature of 739 degrees Celsius (1362 degrees 55 Fahrenheit).

EXAMPLE II

An FCC reaction zone continually receives the regenerated catalyst from the regeneration zone of Example I while sending coke contaminated catalyst to the regenerator in the amount previously stated. The reaction zone, except for the influence of catalyst loadings is operated at a pressure of 206 k pag. A total of 6,470 cubic meters per day of a vacuum gas oil feed is charged to the FCC riser at a temperature of about 262 degrees Celsius (504 degrees Fahrenheit). Properties of the charge stock are given in Table 1.

TABLE 1

FEEDSTOCK PROPERTIES	
API	21.20
PCT Sulfur	1.96 wt. %
Vanadium, PPM	.50
Nickel, PPM	.28
Conradson Carbon Content	.36 wt. %
IBP	343° C.
10%	382° C.
50%	441° C.
90%	498° C.
E.B.	552° C.

The feedstock is contacted with 1,354,634 kg per hour of regenerated catalyst at a temperature of 739 degrees Celsius (1362 degrees Fahrenheit) in the lower portion of the reaction zone riser. After contact in the riser for approximately 2-5 seconds, a catalyst and hydrocarbon vapor mixture having an average temperature of 525 degrees Celsius (977 degrees Fahrenheit), is separated in a disengagement zone of the reactor. Adsorbed hydrocarbon vapors are stripped from downwardly descending catalyst by countercurrent contact with steam. The stripped catalyst particles having an average temperature of 525 degrees Celsius (977 degrees Fahrenheit) enter the regeneration zone of Example I with the properties described therein. The composition of the total reactor product is summarized in Table 2.

TABLE 2

REACTOR PRODUCT COMPOSITION	
	Wt. %
H ₂ S	.76
C ₂ Minus	4.36
C ₃	5.95
C ₄	9.44
Gasoline	42.84
IBP	46.0° C.
EP	221.0° C.
Light Cycle Oil	17.2
IBP	215.0° C.
EP	338.0° C.
Clarified Oil	14.85
Coke	4.59
Total	99.99

EXAMPLE III

The operation of a regeneration zone modified by the method of this invention is described herein. This regeneration zone is operated to remove coke deposits from spent catalyst used in a reaction zone processing the same quantity of feed as that given in Example II. In this example 1,803,710 kg per hour of spent catalyst containing 0.77 wt. % coke enters a regeneration zone as shown in FIG. 2. Thus a total of 13,888 kg per hour of coke enter the regeneration zone. The temperature of the spent catalyst entering mixing riser 22 is again 525 degrees Celsius (977 degrees Fahrenheit). Apart from the influence of catalyst loadings the regeneration zone is operated at 227 k pag. Catalyst from dense bed regeneration zone 32 is transferred via standpipe 21' at a rate of 1,803,693 kg per hour to mixing riser 22 where it is mixed with the spent catalyst and 118,866 kg per hour air having a temperature of 163 degrees Celsius (325 degrees Fahrenheit). The density of the catalyst and gas catalyst mixture again varies from 48-320 kg/m³ in the lower combustion zone and from 16-80 kg/m³ in the combustion riser zone.

Catalyst and regeneration gases are initially separated upon discharge from the combustion riser 33. Separation causes catalyst particles to move downward and collect in dense bed regeneration zone 32 along with catalyst from cyclones 34. The dense bed regeneration zone is operated as a dense fluidized bed having an average density of 320-801 kg/m³. Air having an initial temperature of 163 degrees Celsius (325 degrees Fahrenheit) is injected into bed 32 at a rate of 40,846 kg/hr. After combination regeneration gas from the riser and dense bed regeneration zone is separated from entrained catalyst in cyclones 34. The combined regeneration stream now relatively free of catalyst has a temperature of 704 degrees Celsius (1300 degrees Fahrenheit) along with a CO₂/CO ratio of 3 and an oxygen content of 0.10 mole percent. Fully regenerated catalyst in bed 32 contains less than 0.05 wt. % carbon but has only been heated to 699 degrees Celsius (1290 degrees Fahrenheit) by the regeneration process.

EXAMPLE IV

The benefits to the reaction zone from the reduced regenerated catalyst temperature are demonstrated in the product yield from the reactor receiving and supplying the catalyst in Example III.

Regenerated catalyst from the regeneration zone of Example III is continually supplied to an FCC reaction zone similar to that described in Example II. The reaction zone is operated at a dilute phase pressure of 206 k pag. Feed totaling 6,470 cubic meters per day of vacuum gas oil enters the lower portion of the riser at a temperature of 262 degrees Celsius (504 degrees Fahrenheit). The properties of the feed are the same as those given in Table 1 of Example II. Regenerated catalyst at a temperature of 699 degrees Celsius (1290 degrees Fahrenheit) contacts the combined feed in the lower portion of the riser at a rate of 1,803,693 kg per hour thereby providing a catalyst to oil ratio of 7.23. After contact in the riser for approximately 2-5 seconds, oil vapor and catalyst are separated in the disengagement zone of the riser. Temperatures of the catalyst and vapor leaving the disengagement zone average 525 degrees Celsius (977 degrees Fahrenheit). Following stripping of hydrocarbon vapors, spent catalyst having 0.77 wt. % coke is returned to regeneration zone III at a temperature of 525 degrees Celsius. A hydrocarbon product stream having the properties listed in Table 3 is recovered from the reaction zone.

TABLE 3
REACTOR PRODUCT COMPOSITION

	Wt. %
H ₂ S	.83
C ₂ Minus	3.44
C ₃	6.47
C ₄	10.33
Gasoline	46.3
IPB	46° C.
EP	221° C.
Light Cycle Oil	15.20
IBP	215° C.
EP	338° C.
Clarified Oil	11.87
Coke	5.56
Total	100.0

A comparison of the foregoing examples illustrates the increased coke burning capacity of a regeneration zone operated in accordance with this invention.

In Example I a total of 11,514 kg per hour of coke is removed from the catalyst. In Example III 13,883 kg per hour of coke is removed to completely regenerate the catalyst and the average temperature of the regenerated catalyst is lowered by 40 Celsius degrees.

As demonstrated by the product compositions from the reaction zones, the lowering of the catalyst temperature and concomitant increase in catalyst to oil ratio improves the yield of valuable products. With the modified regeneration zone the same reactor operations convert the same amount of feed into an additional 3.5 wt. % of gasoline in spite of the higher coke yield for the lower temperature operation. This increase in gasoline yield results in only a 2 wt. % loss in light cycle oil. The examples also show an overall reduction in the air requirements for the regeneration zone of this invention. In the conventional regeneration zone of Example I, 14.2 kg of air were required for each pound of coke removed whereas in Example III the regeneration zone of the invention required only 11.5 kg of air per pound of coke. Thus the same amount of feed can be processed more advantageously as a result of this regeneration process.

The increased coke burning capacities of this invention are established by the preceding examples. It is also readily appreciated from the examples that the herein described regeneration method will permit larger quantities of feed, or feeds having increased coke producing tendencies, to be processed at optimum reaction conditions.

It is believed that the benefits achieved by increasing the amount of highly regenerated catalyst which can be delivered to the reaction will also transfer to other hydrocarbon conversion processes which may employ a fluidized reactor.

We claim as our invention:

1. A process for the regeneration of spent hydrocarbon conversion catalyst withdrawn from a fluidized reaction zone, which process comprises the steps of:

(a) passing to a lower locus of a combustion zone of a riser-type fluidized regeneration zone: (i) spent catalyst from said reaction zone, (ii) a stream comprising regenerated catalyst from a hereinafter described dense bed regeneration zone, and (iii) a first oxygen containing regeneration gas stream in an amount sufficient to maintain fast fluidized conditions, which include a catalyst density of 3 to 25 lbs. per cubic foot and a superficial gas velocity of 3 to 10 feet per second, within said combustion zone and to oxidize coke along with coke combustion by-products;

(b) oxidizing coke and coke combustion by-products in said combustion zone while transporting said spent and regenerated catalyst upward in cocurrent flow with rising regeneration gas;

(c) passing catalyst and regeneration gas upward in cocurrent flow from the combustion zone into a riser regeneration zone located above said combustion zone and having a catalyst density of from 1 to 8 pounds per cubic foot and lower than said combustion zone and a gas velocity greater than 10 feet per second, and therein oxidizing coke and coke combustion by-products to produce partially regenerated catalyst and a spent first regeneration gas, having a mole ratio of CO₂/CO of from about 0.7 to about 2.0;

(d) discharging partially regenerated and regenerated catalyst and said spent first regeneration gas from

an upper locus of said riser regeneration zone into a catalyst disengagement zone through an outlet means that effects at least a partial separation of catalyst and regeneration gas and thereby causing an initial separation of catalyst and the spent first regeneration gas;

- (e) allowing partially regenerated and regenerated catalyst discharged through said outlet means to settle downward through a dilute phase above a dense fluidized bed, having a density of 30 to 50 pounds per cubic foot, of a dense bed regeneration zone located below said dilute phase and said disengagement zone while flowing countercurrent to a rising second regeneration gas, and introducing into the dense fluidized bed a second oxygen containing regeneration gas stream in a quantity at least sufficient to produce regenerated catalyst having less than 0.1 wt. % coke and to oxidize essentially all of the carbon monoxide produced within said dense bed regeneration zone to carbon dioxide;
- (f) combining said spent first regeneration gas with the second regeneration gas, which contains at least 0.5 mole % O₂ and emanates from said dense bed regeneration zone in said disengagement zone, 25 and producing a combined spent regeneration gas having a CO₂/CO mole ratio in the range of about 1 to 5;
- (g) withdrawing said combined spent regeneration gas from the disengagement zone; and,
- (h) withdrawing regenerated catalyst from said dense bed having an average temperature in the range of 620 to 787 degrees Celsius and returning at least a first portion of said regenerated catalyst to said reaction zone and at least a second portion of regenerated catalyst to the lower locus of the combustion zone pursuant to step (a) above.

2. The process of claim 1 wherein the quantity of regeneration gas added to said dense bed regeneration zone is sufficient to provide oxygen required to oxidize carbon monoxide in said spent regeneration gas discharged from said riser regeneration zone in addition to producing regenerated catalyst within the dense bed regeneration zone.

3. The process of claim 1 wherein regenerated catalyst is returned from said dense bed regeneration zone to said combustion zone through an internal standpipe.

4. The process of claim 1 wherein spent catalyst, the regenerated catalyst stream, and regeneration gas are combined in a mixing zone prior to entering said combustion zone.

5. The process of claim 1 wherein between 65 and 90 wt. percent of the coke on the spent catalyst entering the combustion zone is removed during passage through the combustion and riser regenerator zones.

6. The process of claim 1 wherein the partially regenerated catalyst and spent first regeneration gas from the riser regeneration zone are transferred to the disengagement zone by a substantially horizontal conduit and the combustion zone and riser regeneration zone are in side-by-side relationship to the dense bed regeneration zone and disengagement zone.

7. A process for the regeneration of particulate hydrocarbon cracking catalyst withdrawn from a fluidized catalytic cracking reaction zone, which process comprises the steps of:

- (a) passing spent catalyst from said reaction zone into a lower locus of a combustor zone of a riser-type

fluidized catalytic cracking regeneration zone together with a stream consisting of regenerated catalyst from a hereinafter described dense bed regeneration zone and an oxygen-containing regeneration gas in sufficient quantity to obtain fast fluidized conditions through said combustor zone and to oxidize coke and coke combustion by-products, said fast fluidized conditions including a catalyst density of 3 to 25 pounds per cubic foot and a superficial gas velocity of 3 to 10 feet per second;

- (b) oxidizing coke and coke combustion by-products in said combustor zone while transporting catalyst upward with concurrently flowing regeneration gas;
- (c) passing said catalyst and regeneration gas mixture upward from the combustor zone into a riser regeneration zone located above said combustor zone and operated at a catalyst density of from 1 to 8 pounds per cubic foot lower than said combustor zone and with a gas velocity greater than 10 feet per second and therein oxidizing coke and coke combustion by-products to produce partially regenerated catalyst and a spent first regeneration gas having a mole ratio of CO₂/CO of from about 0.07 to 2.0;
- (d) discharging fully regenerated and partially regenerated catalyst and the spent first regeneration gas from an upper locus of said riser regeneration zone into a catalyst disengagement zone through an outlet means that effects at least a partial separation of catalyst and regeneration gas and thereby causing an initial separation of partially regenerated catalyst and said first regeneration gas;
- (e) allowing at least 70 wt. % of the partially regenerated catalyst discharged into the catalyst disengagement zone through said outlet means to settle downward through a dilute phase located above a dense phase catalyst bed having a density of 30 to 50 pounds per cubic foot retained in a dense bed regeneration zone located below said dilute phase and said disengagement zone, and introducing into the dense bed regeneration zone a second oxygen-containing regeneration gas stream at a rate which is adjusted in response to a hereinafter defined output signal;
- (f) combining said spent first regeneration gas with a second regeneration gas, in said disengagement zone to obtain a combined regeneration gas having a CO₂/CO mole ratio of from 1 to about 5, said second regeneration gas comprising at least 0.5 mole % oxygen and emanating from said dense bed regeneration zone and rising upward countercurrent to descending catalyst, and removing the combined regeneration gas from said disengagement zone;
- (g) measuring the temperature of the combined regeneration gas, comparing said temperature to a set point, and generating from this comparison said output signal, which determines the flow rate of the second oxygen-containing regeneration gas stream to the dense bed regeneration zone such that said set point is not exceeded; and
- (h) withdrawing regenerated catalyst from said dense bed regeneration zone having an average temperature in the range of 620 to 787 degrees Celsius and returning said regenerated catalyst particles to said reaction zone.

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8. The process of claim 7 wherein said set point is equal to a maximum operating temperature for the regeneration equipment.

9. The process of claim 8 wherein the set point is equal to 815 degrees Celsius.

10. The process of claim 7 wherein said set point is a temperature measured within the disengagement zone,

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and the comparison provides a measure of the degree of afterburning occurring within the disengagement zone.

11. The process of claim 7 wherein at least 80 wt. percent of the catalyst discharged from the upper locus of said riser regeneration zone falls downward countercurrent to rising gas having an oxygen content above 0.5 mole percent.

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Claim 5 C/3

United States Patent

[19]

Stine et al.

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[11] 3,919,115

[45] *Nov. 11, 1975

[54] FLUIDIZED CATALYST REGENERATION PROCESS

[75] Inventors: Laurence O. Stine, Western Springs; Algie J. Conner, Downers Grove, both of Ill.

[73] Assignee: Universal Oil Products Company, Des Plaines, Ill.

[*] Notice: The portion of the term of this patent subsequent to Oct. 29, 1991, has been disclaimed.

[22] Filed: Mar. 25, 1974

[21] Appl. No.: 454,629

Related U.S. Application Data

[63] Continuation of Ser. No. 263,399, June 8, 1972, abandoned.

[52] U.S. Cl. 252/417; 208/120; 208/164; 252/419

[51] Int. Cl. B01J 37/12; C10G 35/00

[58] Field of Search..... 252/417, 419, 414; 208/164, 120; 23/288 B, 288 S

[56] References Cited

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[57] ABSTRACT

A process conducted in a regeneration zone, comprising a first dense bed, a dilute phase transport riser, and a second dense bed, wherein coke-contaminated spent catalyst is oxidized to a controllable residual coke level and CO thereby produced is essentially completely burned to CO₂. Spent catalyst is regenerated in the first dense bed to the desired coke level by controlling the stream of fresh regeneration gas to that bed. Combustion of CO takes place in the dilute phase transport riser through which regenerated catalyst passes at residence times sufficiently short to preclude further regeneration.

5 Claims, 3 Drawing Figures

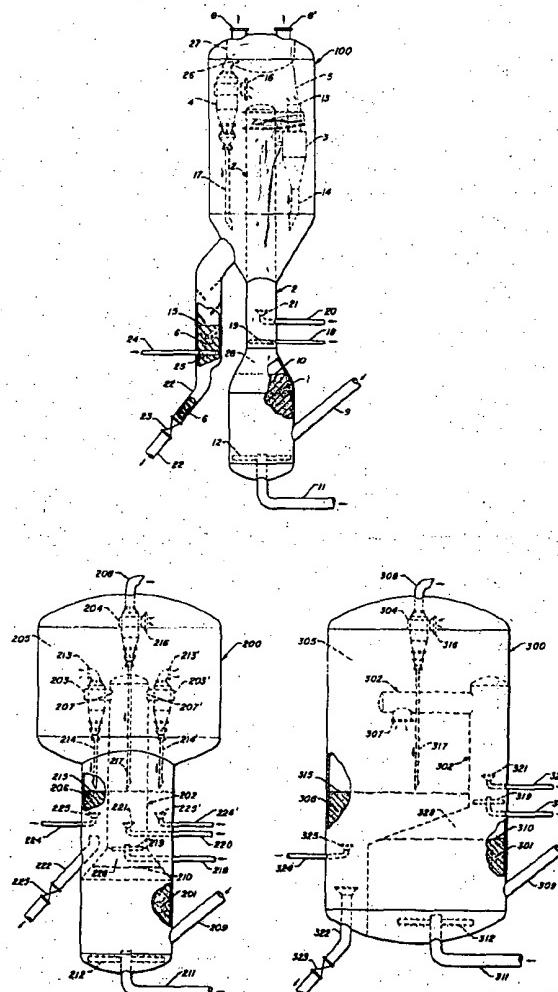
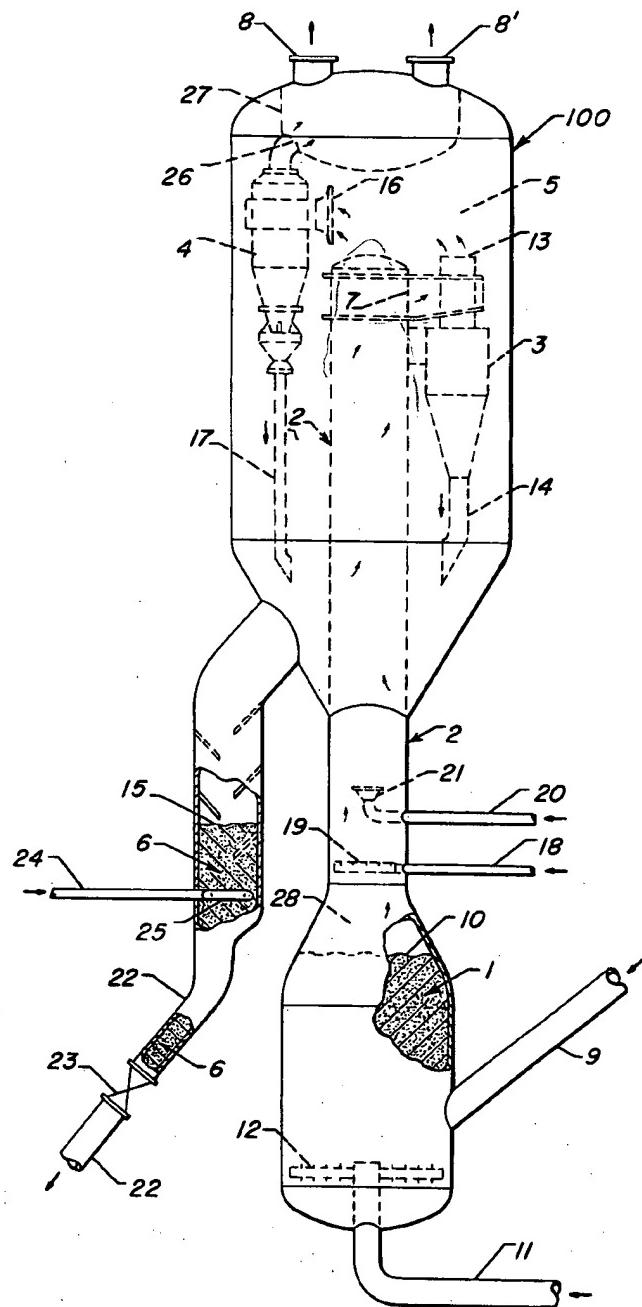
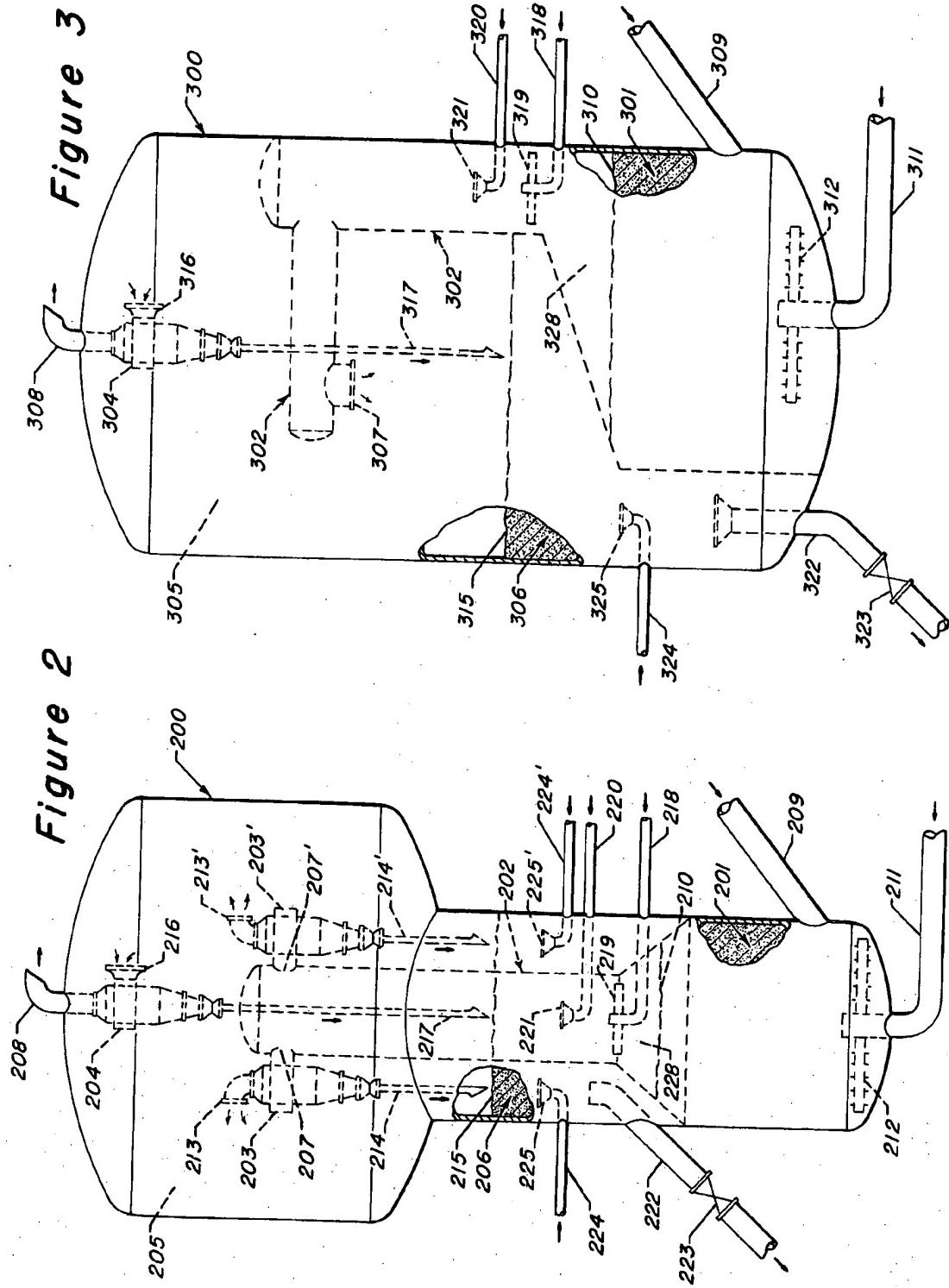


Figure 1





FLUIDIZED CATALYST REGENERATION PROCESS

CROSS-REFERENCES TO RELATED APPLICATION

This is a continuation application of pending application Ser. No. 263,399, filed on June 8, 1972, now abandoned.

By controlling the residual coke level on regenerated catalyst, the process allows the refiner to directly and conveniently control the conversion level and product distribution obtained in the reaction zone. At the same time within the regeneration zone combustion of CO substantially eliminates CO pollution of the atmosphere and recovers the heat of combustion without the need for a CO boiler which decreases feed preheat requirements. Higher velocities and higher O₂ concentrations employed in the process result in shorter catalyst and gas residence times, reduced catalyst inventory, lower catalyst makeup rate, improved regeneration, and improved catalyst stability.

BACKGROUND OF THE INVENTION

1. Field of the Invention

The field of art to which the invention pertains is hydrocarbon processing. More particularly, this invention relates to systems in which a fluidized catalyst is continuously recycled between a reaction zone wherein it becomes spent by carbonaceous deposits and a separate regeneration zone wherein its activity is restored by removal of the deposits. Specifically, the present invention relates to a regeneration process which is applicable for the controlled oxidative removal of carbonaceous deposits from a spent fluidized catalytic cracking catalyst along with the essentially complete combustion of CO thereby produced.

2. Description of the Prior Art

There are a number of continuous cyclical processes employing fluidized solid techniques in which carbonaceous materials are deposited on the solids in the reaction zone and the solids are conveyed during the course of the cycle to another zone where carbon deposits are at least partially removed by combustion in an oxygen-containing medium. The solids from the latter zone are subsequently withdrawn and reintroduced in whole or in part to the reaction zone. Among such processes are fluid coking, fluid dehydrogenation, and fluid catalytic cracking.

One of the more important processes of this nature is the fluid catalytic cracking process for the conversion of relatively high-boiling hydrocarbons to lighter hydrocarbons boiling in the heating oil and gasoline (or lighter) range. The hydrocarbon feed is contacted in one or more reaction zones with the particulate cracking catalyst maintained in a fluidized state under conditions suitable for the conversion of hydrocarbons.

The gaseous effluent from the reaction zone is passed to a product recovery zone while the catalyst is generally passed to a stripping zone for removal of strippable hydrocarbons from the particles. The stripped catalyst is subsequently introduced into a fluidized regeneration zone where non-strippable carbonaceous material is contacted with an oxygen-containing gas, for example air, under conditions such that a major portion of the carbon on the catalyst particles is removed therefrom by combustion.

Generally, the regeneration is done in a single dense bed located in the bottom portion of the regeneration zone. Provisions are made for recovering and returning to the dense bed catalyst entrained in the flue gas effluent passing from the dense bed. This is generally carried out by passing this effluent flue gas containing entrained catalyst through cyclones located in the rather large disengaging space which is positioned above and in connection with the dense bed.

10 Superficial velocities within the regeneration zone are generally within the range of about 1.5 to 6 feet per second with 1.5 to 3 feet per second being the more common range. Residence time of the catalyst within the regeneration zone is generally in the 2 to 5 minute 15 range with 2 to 3 being the more common, while the residence time of gas is generally within the range of 10 to 20 seconds.

It is the present practice to control the oxygen-containing gas stream introduced to such regeneration 20 zone directly responsive to a predetermined temperature differential between the gas outlet disengaging space and the dense bed of the regeneration zone to minimize excess oxygen therein and to thereby control and limit the amount of afterburning of CO to CO₂ in 25 the upper disengaging portion of the regeneration zone. This restricting of afterburning of CO is done for the very practical reason of avoiding the damaging effects of excessively high temperatures in the upper disengaging space. This practice, as exemplified by Pohlenz U.S. 30 Pat. Nos. 3,161,583 and 3,206,393, produces a small amount of oxygen in the flue gas, generally in the range of 0.1 to 1% oxygen.

More importantly, the present practice of controlling the amount of fresh regeneration gas such as to permit 35 a slight amount of afterburning also fixes the degree of catalyst regeneration, that is, the residual coke on the regenerated catalyst. Although it is widely known that the level of coke content on regenerated catalyst has a great influence on the conversion and product distribution 40 obtained in the reaction zone, residual coke level on regenerated catalyst produced by present regeneration processes is not an independent variable. The coke level when the controlled afterburning technique is used is fixed by regenerator design at some low level, typically from about 0.05 to 0.4 wt. % carbon, and more often from about 0.15 to 0.35 wt. % carbon.

Present industry practice is generally to direct the flue gas which contains CO either directly to the atmosphere or to a CO boiler where it is used as fuel to make steam. In another application, as disclosed in Campbell U.S. Pat. No. 3,363,993, the flue gas is fired as fuel in a heater which is used to preheat the fresh feed charged to the reaction zone of the fluid catalytic cracking process. Pfeiffer et al, U.S. Pat. No. 3,104,227, discloses 50 directing the flue gas to an expanderturbine CO boiler combination to produce the power required to drive the fresh regeneration gas compressor.

Although the efficiency of cyclones and other such equipment for the recovery of solid catalyst particles is 55 usually very high, some catalyst is always physically lost from the regeneration zone. With time, catalyst within the system loses activity and therefore becomes effectively lost because of the cumulative effects of exposure to contaminant metals contained in the fresh feed, high temperatures, and steam. As well, equilibrium catalyst must be withdrawn and replaced with fresh catalyst to maintain desired catalyst activity. For these reasons, it is necessary or desirable to add fresh make-up

catalyst to maintain the desired total inventory at the desired equilibrium activity of the catalyst. Typical daily make-up catalyst rates are between about 0.5 to 2.0% of total catalyst inventory in the system with the average being about 1% of inventory per day. Because fluid catalytic cracking catalysts, particularly those containing zeolites, are relatively expensive, it is evident that processes requiring large inventories and therefore large catalyst make-up rates tend to be less economically attractive. As a result thereof, every effort is made to reduce the initial catalyst inventory investment and that portion of overall operating costs associated with catalyst make-up.

Since most of the catalyst inventory is contained within the regeneration zone, it is the present practice to employ operating conditions in the regeneration zone that favor high carbon burning rates, thereby permitting lower regenerator inventories. Such desired operating conditions are high partial pressure of oxygen and high temperature. There have been, therefore, recent industry trends toward higher pressure and higher temperature regenerators for this reason of obtaining higher burning rates. Previously the preferred pressure range had been from about 10 to about 25 psig and preferred temperatures had been in the 1100°-1150°F range. Pressures in the range of 30-40 psig and temperatures in the range of 1150° to 1250°F or higher are now rather common. Although some inventory reduction has been achieved, limitations on these approaches have been imposed by higher equipment costs due to higher pressures and by increased catalyst deactivation due to long residence time in the high temperature regenerators, and by the terminal velocity above which the catalyst cannot be maintained in the lower part of the regeneration zone as a dense bed.

Further attempts to reduce catalyst inventories and makeup rates have been made by effecting a staged regeneration within a regeneration vessel. U.S. Pat. Nos. 3,494,858 (E. C. Luckenbach) and 3,563,911 (R. W. Pfeiffer and L. W. Garret, Jr.) are such examples.

U.S. Pat. No. 3,494,858 discloses a countercurrent regeneration process in which spent catalyst is partially regenerated in a first fluidized bed with partially spent regeneration gas, further regenerated in a transfer line regeneration zone with fresh regeneration gas and then passed to a second fluidized bed wherein further regeneration may or may not take place with partially spent regeneration gas. Preferred superficial velocities are in the range of 2.5 to 3.0 ft./sec. and preferred temperatures are about 1100° to 1175°F. No afterburning of CO to CO₂ is mentioned but "substantially complete" removal of carbon to a level below 0.2% and possibly as low as 0.1% is contemplated. Substantially completely regenerated catalyst may then be stripped of high oxygen-containing gases in a separate stripping zone with steam or flue gas.

U.S. Pat. No. 3,563,911 discloses a two-stage regeneration process in which spent catalyst is partially regenerated in a first dense bed with a first oxygen-containing gas stream and then further regenerated in a second dense bed with a second oxygen-containing gas stream. A common dilute phase is superimposed above both dense beds. Preferably, superficial velocities are maintained in the range of about 2 to 4.5 feet per second and the preferred temperature range is from about 1125°F to 1350°F. It is desirable in the process of this invention to control by means of a flue gas and last bed temperature differential the amount of oxygen-contain-

ing gas admitted to the regeneration zone such that only a small amount of CO afterburning takes place. This is consistent with present industry practice on single-stage regeneration zones. Examples are presented which indicate that at gas velocities of 2.5 and 4.5 feet per second some degree of inventory reduction over that of a particular type of single stage regeneration can be achieved by staging. A disadvantage, however, is that there is an upper limit of residual coke to which the catalyst can be regenerated by staging with beneficial results.

The process of our invention provides not just for limited, controlled afterburning of CO but for the essentially complete combustion of the CO present. For this purpose, and others, a specific dilute phase transport riser has been provided. By so separating CO oxidation from coke oxidation within the regeneration zone, fresh regeneration gas can now be controlled to obtain the degree of coke oxidation desired rather than to limit the amount of CO afterburning. By the process of our invention, therefore, residual coke on regenerated catalyst can be controlled directly for the purpose of controlling reaction zone conversion and product yield structure. At the same time, the CO so produced from coke oxidation is essentially completely combusted, thereby eliminating an air pollution problem without the need for a CO boiler at the same reducing feed preheat requirements. Further, by employing in the process of our invention higher velocities and temperatures than those presently used, dramatic catalyst inventory reductions, independent of residual coke level, are possible. Additional advantages of higher velocities and reduced inventory are shorter catalyst residence times, lower catalyst makeup rates, improved regeneration and improved catalyst stability.

SUMMARY OF THE INVENTION

Our invention can briefly be summarized as a process for the regeneration of coke contaminated catalyst, removed from a reaction zone, and the combustion of CO thereby produced wherein the residual coke on regenerated catalyst is directly controllable. A broad embodiment of our invention resides in the process comprising the steps of: (a) passing said catalyst to a first dense bed of particulate material in a regeneration zone and controlling the rate of fresh regeneration gas, to said first dense bed, to oxidize said coke to a predetermined residual coke level; (b) oxidizing coke in said first dense bed at oxidizing conditions to produce partially spent regeneration gas, containing CO, and regenerated catalyst having residual coke thereon; (c) passing said regenerated catalyst and partially spent regeneration gas to a dilute phase transport riser, in said regeneration zone, and therein converting CO to CO₂ at conversion conditions, to produce spent regeneration gas; (d) separating the thus-regenerated catalyst from said spent regeneration gas; and, (e) introducing said regenerated catalyst into a second dense bed of particulate material from which said regenerated catalyst is returned to said reaction zone.

DESCRIPTION OF THE DRAWING

Having thus described the invention in brief general terms, reference is now made to the schematic drawings in order to provide a better understanding of the present invention.

It is to be understood that the drawings are shown only in such details as are necessary for an understand-

ing of the invention and that various items such as valves, bleed and dispersion steam lines, instrumentation and other process equipment and control means have been omitted therefrom for the sake of simplicity.

The drawings of this specification include FIG. 1 which depicts schematically a specific apparatus suitable for carrying out the process of our invention, and FIG. 2 and FIG. 3 which depict alternate apparatus also suitable for carrying out this invention. FIGS. 2 and 3 include the same principles as FIG. 1; FIG. 3 is of particular interest because it indicates how the present invention might be applied to an existing regeneration vessel.

FIG. 1 shows a regeneration apparatus 100 basically containing a first dense bed 1, a dilute phase transport riser 2, a catalyst and regeneration gas separation means 3 and 4, a disengaging space 5 and a second dense bed 6. First dense bed 1 is shown at the lower portion of the figure and is connected to one of a vertically extending dilute phase riser 2 through a transition region 28. Catalyst and regeneration gas separation means 3 and 3' are attached to the outlets 7 of transport riser 2. Separated regeneration gas leaving separation means 3 passes into disengaging space 5 then into separation means 4, exits at 26 into plenum chamber 27. Regeneration gas leaves the plenum chamber 27 and the regeneration zone via outlets 8 and 8'. Separated catalyst from separation means 3 and 3' and 4 is directed to second dense bed 6.

Spent catalyst is introduced through inlet line 9 to the first dense bed 1 having a level indicated at 10 which is located in a transition region 28 positioned between the first dense bed 1 and the transport riser 2. A controlled stream of fresh regeneration gas is introduced via line 11 into dense bed 1 through distributing device 12 which allows the fresh regeneration gas to be more readily dispersed within the dense bed 1. Typically the distributing device can be a metal plate containing holes or slots or preferably a pipe grid arrangement, both types of which are quite familiar to those skilled in the art. Controlled oxidation of the carbonaceous deposits to the desired residual carbon level takes place in dense bed 1 and partially spent regeneration gas and fluidized regenerated catalyst are carried out of bed 1 through the transition region 28 into transport riser 2 wherein CO oxidation takes place to produce spent regeneration gas and where the heat of combustion of the CO to CO₂ is transferred to the catalyst being transported.

Transport riser 2 is vertically positioned having its inlet at the lower portion and its outlet means 7 near its top portion. Outlet means 7 may be single or multiple openings located at or near the uppermost portion of transport riser 2 which allow regenerated catalyst and spent regeneration gas to pass out of transport riser 2. As shown in FIG. 1 catalyst and regeneration gas separation means 3 is attached to the outlets 7 of transport riser 2. Separation means 3, typically cyclone separation means, are used to achieve a substantial separation of spent regeneration gas and entrained catalyst which pass out of transport riser 2. Although the FIG. 1 shows only one such cyclone attached to the transport riser 2, it is contemplated that from 1 to 4 cyclones could be so positioned. Entrained catalyst and regeneration gas pass via outlet 7 into the separation means 3 out of which spent regeneration gas substantially free of catalyst passes out of the separation means through outlet 13 while catalyst passes through dipleg 14 directed

toward second dense bed 6. Optionally, separation means 3 could be omitted, leaving regeneration gas and catalyst exiting outlet means 7 to directly enter disengaging space 5. Some separation of regeneration gas and catalyst would be achieved but not as efficiently as with cyclone separation means 3 as shown.

Separation means 4, also typically a cyclone separation means, has an inlet 16 which receives regeneration gas and any entrained catalyst located in the disengaging space 5. Regeneration gas and any entrained catalyst are substantially separated from each other with the spent regeneration gas passing out of the separation means 4 at outlet 26, and into plenum chamber 27 and then out of the regeneration zone via spent regeneration gas outlet 8 and 8'. Catalyst separated from the regeneration gas is passed via dipleg 17 downward toward second dense bed 6 having a level indicated at 15.

An external combustible fluid such as fuel gas or a liquid hydrocarbon stream is admitted to the transport riser 2 through line 20 via distributor 21. The burning of such a fluid may be necessary to assist in initial process startup, to increase the temperature within the dilute phase transport zone 2 sufficiently to initiate CO oxidation, and may be required to further increase the temperature of the catalyst particles passing through the riser, beyond that which could be achieved by the CO burning alone. The temperature of the regenerated catalyst delivered to the reaction zone can, therefore, be controlled independently of the degree of coke oxidation by controlling the stream of combustible fluid burning in the transport riser. Although not shown in the Figures, additional fuel could be added to the first dense bed 1 for some of or all the above reasons.

Additionally, a second stream of fresh regeneration gas is admitted to the transport riser through line 18 via distributor 19. This fresh regeneration gas stream may be for the purpose of supplying needed oxygen to support burning of the external combustible gas or may be done for the purpose of venting fresh regeneration gas in excess of that required for the degree of coke oxidation desired in dense bed 1. As well, this stream of regeneration gas could be used to cool the regenerated catalyst down to the predetermined catalyst delivery temperature in the event that the complete combustion of CO produced higher than desired catalyst temperature. Although one inlet line for the second stream of regeneration gas is shown, there could be multiple inlet lines positioned along the length of the transport riser.

Catalyst passing through diplegs 14 and 17 discharges downward toward the second dense bed 6.

This second dense bed 6 is so positioned in relationship to the first dense bed 1 as to maintain a sufficient head of regenerated catalyst necessary to overcome any pressure drop in regenerated catalyst exit line 22, control valve 23, and in any equipment downstream connected to line 22. Although FIG. 1 shows the second dense bed 6 positioned immediately above the first dense bed 1, they may be separated by a sufficient distance to provide this head. Furthermore, the level 15 of catalyst within the second dense bed 6 may be controlled as to provide the desired residence time within the bed. It should be noted that second dense bed 6 need not be a stripper as indicated in FIGS. 1, 2 and 3. The term second dense bed as used in this specification means regenerated catalyst maintained in dense bed for head and sealing purposes prior to being returned to the reaction zone.

The catalyst in bed 6 moves in a downward direction and eventually passes out of the regeneration vessel via conduit 22. Also located on conduit 22 is a valve 23 which may be used to control the rate of withdrawal of regenerated catalyst from bed 6. Typically, valve 23 is a slide valve and is operated by a reactor temperature controller or level controller.

A stripping medium can be admitted to the second dense bed 6 through line 24 via distributor 25 to strip from the regenerated catalyst adsorbed and interstitial regeneration gas. Generally the stripping medium will be superheated steam.

It is anticipated in the process of this invention that most of the catalyst within the regeneration zone will be contained in the first dense bed with the smaller portion contained in the second dense bed. More specifically when steam stripping is employed within the second dense bed 6, the second dense bed volume will be so designed such that the catalyst residence time within the bed will be less than 1 minute and preferably less than 30 seconds.

Shown in FIG. 2 is an alternate apparatus 200 in which this invention may be practiced. Primary components are: a first dense bed 201, a dilute phase transport riser 202, a transition region 228, catalyst and regeneration gas separation means 203 and 203' and 204, a disengaging space 205 and a second dense bed 206. First dense bed is again shown at the lower portion of the figure and is connected to one end of a vertically extending dilute phase riser 202 through a transition region 228. Catalyst and regeneration gas separation means 203 and 203' are attached to outlets 207 and 207' of transport riser 202. Separated regeneration gas leaving separation means 203 and 203' via outlets 213 and 213' passes into disengaging space 205, then into separation means 204, and finally exits the regeneration zone via line 208. Separated catalyst from separation means 203 and 203' and 204 is directed via diplegs 214, 214' and 217 to second dense bed 6.

Spent catalyst is introduced through inlet line 209 to the first dense bed 201 having a level indicated at 210 which is within the transition region 228. A controlled stream of fresh regeneration gas is introduced via line 211 into dense bed through distributing device 212 which allows the fresh regeneration gas to be more readily dispersed within the dense bed 201. Typically the distributing device can be a metal plate containing holes or slots or preferably a pipe grid arrangement, both types of which are quite familiar to those skilled in the art. Oxidation of the carbonaceous deposits to a residual carbon level, which can be independently controlled if desired, takes place in dense bed 201 and partially spent regeneration gas and fluidized regenerated catalyst are carried out of bed 201 through the transition region 228 and into transport riser 202 wherein CO oxidation takes place to produce spent regeneration gas and where heat is transferred to the catalyst.

Transport riser 202 is vertically positioned having its inlet at the lower portion and its outlet means 207 and 207' near its top portion. Outlet means 207 and 207' may be openings located at or near the uppermost portion of transport riser 202 which allow catalyst and regeneration gas to pass out of transport riser 202. As shown in FIG. 2 catalyst and regeneration gas separation means 203 and 203', typically cyclones, are attached to the outlets 207 and 207' of transport riser 202 and are used to achieve a substantial separation of spent regeneration gas and entrained catalyst passing

out of transport riser 202 so that the material present in the disengaging space 205 is essentially regeneration gas with very little entrained catalyst present. Entrained catalyst and regeneration gas pass from riser 202 via outlets 207 and 207' into the separation means 203 and 203' where regeneration gas substantially free of catalyst passes out of the separation means through outlets 213 and 213' and catalyst passes through diplegs 214 and 214' downward toward second dense bed 206 having a level indicated at 215. As mentioned in the FIG. 1 description, separation means 203 and 203' could be omitted, leaving regeneration gas and catalyst exiting outlet means 207 and 207' to directly enter disengaging space 205. Some separation of regeneration gas and catalyst would be achieved but not as efficiently as with cyclone separation means 203 and 203' as shown.

Separation means 204, also typically a cyclone separation means, has an inlet 216 which receives regeneration gas and any entrained catalyst located in the disengaging space 205. Regeneration gas and any entrained catalyst are substantially separated from each other with the regeneration gas passing out of the separation means 204 and out of the regeneration zone via regeneration gas outlet 208. Catalyst separated from the regeneration gas is passed via dipleg 217 downward toward the second dense bed 206.

An external combustible fluid such as fuel gas or a liquid hydrocarbon stream is admitted to the transport riser 202 through line 220 via distributor 221. The burning of such a fluid may be necessary, as previously mentioned, to assist in initial process startup, to increase the temperature within the dilute phase transport riser 202 sufficiently to initiate CO oxidation, and it may also be required to further increase the temperature of the catalyst particles passing through the riser, beyond that which could be achieved by the CO burning alone. The temperature of the regenerated catalyst delivered to the reaction zone can, therefore, be controlled independently of the degree of coke oxidation by controlling this stream of combustible fluid burning in the transport riser.

Additionally, a second stream of fresh regeneration gas is admitted to the transport riser through line 218 via distributor 219. This fresh regeneration gas stream as mentioned in the FIG. 1 description may be for the purpose of supplying needed oxygen to support burning of the external combustible gas or may be done for the purpose of venting fresh regeneration gas in excess of that required for the degree of coke oxidation desired in dense bed 201. As well, this stream of regeneration gas could be used to cool the regenerated catalyst down to the predetermined catalyst delivery temperature in the event that the complete combustion of CO produced higher than desired catalyst temperature. Although one inlet line for the second stream of regeneration gas is shown in FIG. 2, there could be multiple inlet lines positioned along the length of the transport riser.

Catalyst passing through diplegs 214 and 214' and 217 discharges downward toward second dense bed 206 which has an interface 215 defining the boundary between the disengaging space 205 and the dense bed 206.

This second dense bed 206 is so positioned above the first dense bed 201 to maintain a sufficient head of regenerated catalyst to overcome any pressure drop in regenerated catalyst exit line 222, control valve 223, and in any equipment downstream connected to line

222. Although FIG. 2 shows the second dense bed 206 positioned immediately above the first dense bed 201 they may be separated by a sufficient distance to provide this head. As a consideration independent of dense bed location, the level 215 of the dense bed can be controlled as required to provide the desired catalyst residence time within the bed.

The catalyst in bed 206 moves in a downward direction and eventually passes out of the regeneration vessel via conduit 222. Also located on conduit 222 is a valve 223 which may be used to control the rate of withdrawal of regenerated catalyst from bed 206. Typically, valve 223 is a slide valve and is operated by a reactor temperature controller or level controller.

A stripping medium can be admitted to the second dense bed 206 through lines 224 and 224' via distributors 225 and 225' to strip from the regenerated catalyst adsorbed and interstitial regeneration gas. Generally the stripping medium will be superheated steam.

As previously mentioned, it is anticipated in the process of this invention that most of the catalyst within the regeneration zone will be contained in the first dense bed with the smaller portion contained in the second dense bed. More specifically when steam stripping is employed within the second dense bed 206, the second dense bed volume will be so designed such that the catalyst residence time within the bed is less than about 1 minute and preferably less than 30 seconds.

FIG. 3 shows another alternate apparatus 300 suitable for practicing this invention. Specifically, FIG. 3 represents a possible modification of existing regenerators to produce an apparatus suitable for carrying out the process of this invention. Basically an existing regeneration vessel 303 is modified to provide a first dense bed 301, a dilute phase transport riser 302, a second dense bed 306 and the associated equipment for riser 302 and bed 306. The portions of FIG. 3 generally function the same as those previously described for FIG. 1 and FIG. 2.

Briefly, spent catalyst is introduced through inlet line 309 into a first dense bed of catalyst 301 having a level or interface indicated at 310, located within transition region 328 which is positioned between first dense bed 301 and transport riser 302. A controlled stream of fresh regeneration gas necessary to achieve the desired residual coke level is introduced at line 311 into dense bed 301 through distributing device 312. Oxidation of coke contained on the spent catalyst takes place in dense bed 301 and partially spent regeneration gas and regenerated catalyst are swept out of bed 301 through transition region 328 and into transport riser 302 wherein CO oxidation takes place and where heat is transferred to the catalyst passing through the transport riser.

As described for FIG. 1, a combustible fluid may be admitted to the transport riser 302 through line 320 via distributor 321 and additionally a second stream of fresh regeneration gas may be admitted to the riser through line 318 via distributor 319.

Entrained catalyst and spent regenerated gas then pass out of transport riser 302 via riser outlet means 307 into disengaging space 305. Preferably outlet means 307 is so placed that it ejects entrained catalyst and spent regeneration gas in a downward direction to reduce the extent of catalyst entrained within the disengaging space 305.

Separation means 304, typically a cyclone separation means, has an inlet 316 and receives spent regeneration

gas and any entrained catalyst from the disengaging space 305. Regeneration gas and any entrained catalyst are substantially separated from each other with the spent regeneration gas passing out of the separation means 304 and out of the regeneration zone 303 via outlet 308. Separated catalyst is passed via dipleg 317 downward toward second dense bed 306 which has an interface 315 defining the boundary between the second dense bed 306 and the disengaging space 305.

Catalyst in bed 306 moves in a downward direction and leaves the bed and the regeneration zone through conduit 322. The rate of catalyst withdraw is controlled by valve 323 which is generally operated by a reactor temperature controller.

The stripping medium can be admitted to the second dense bed 325 through line 324 via distributor 325 to effect stripping of regeneration gas from regenerated catalyst.

DESCRIPTION OF THE INVENTION

Catalysts which can be used in the process of this invention include those known to the art as fluidized catalytic cracking catalysts. Specifically the high activity zeolite-containing catalyst can be used and are preferred because of their higher resistance to the deactivating effects of high temperatures, exposure to steam, and exposure to metals contained in the feedstock. The well-known amorphous catalysts may also be used and are especially suited to the process of this invention because the lower residence times employed in this process will extend their useful life.

Charge stocks used in the catalytic cracking process are mentioned here because contaminants such as Conradson carbon and metals such as nickel, iron, and vanadium found in the charge stock usually influence the regeneration operation, catalyst selectivity, catalyst activity and the fresh catalyst makeup rate required to maintain a constant activity. A high carbon content of the feedstock as measured by a Conradson carbon analysis reduces the capacity of those units that are limited by regeneration capacity. Furthermore, regeneration temperatures tend to increase with increasing carbon content. Metals contained in the feed are also deposited on the catalyst and not only change its selectivity in the direction of less gasoline and more coke and light gas in a given reactor system but tend to deactivate the catalyst.

Because of a deleterious effect of Conradson carbon and metals, only feeds substantially free or containing but limited amounts of them are charged to most present cracking operations. For the reduction of Conradson carbons and metals such processes as vacuum distillation, deasphalting, visbreaking and coking have been used to prepare suitable feedstocks for fluid catalytic cracking. Typical feedstocks include the vacuum gas oils which boil within the range of about 400° to about 1000°F or higher depending on the limits set for Conradson carbons and metals. The term feedstocks can also include recycled materials such as gasoline, light or heavy cycle oils or slurry oil which are separated from the hydrocarbon effluent leaving the reaction zone by the main fractionation column. Because of the improved solid gas contact, shorter residence time and smaller catalyst inventories of the process of the present invention, it is an advantage of this process over those employing conventional regeneration techniques that heavier, more contaminated feedstocks can be charged to the reaction zone for a given air rate (or

coke burning capacity) and a given catalyst makeup rate.

This aspect and other aspects and advantages of the present process will be made clearer by a brief reference to a typical fluid catalytic cracking process in present industry use with particular emphasis on the regeneration zone of such process.

In a typical FCC process flow, finely divided regenerated catalyst leaves the regeneration zone at a certain temperature and contacts a feedstock in a lower portion of a reaction riser zone. While the resulting mixture passes up through the riser, conversion of the feed to lighter products and to coke deposited on the catalyst occurs. The effluent from the riser is discharged into a disengaging space where additional conversion can take place. The hydrocarbon vapors, containing entrained catalyst, are then passed through one or more cyclone separation means to separate any spent catalyst from the hydrocarbon vapor stream. The separated hydrocarbon vapor stream is passed into a fractionation zone known in the art as the main column wherein the hydrocarbon effluent is separated into such typical fractions as light gases and gasoline, light cycle oil, heavy cycle oil and slurry oil. Various fractions from the main column can be recycled along with the feedstock to the reaction riser. Typically fractions such as light gases and gasoline are further separated and processed in a gas concentration process located downstream of the main column. Some of the fractions from the main column as well as those recovered from the gas concentration process may be recovered as final product streams. The separated spent catalyst passes into the lower portion of the disengaging space and eventually leaves that zone passing through stripping means in which a stripping gas, usually steam, countercurrently contacts the spent catalyst purging adsorbed and interstitial hydrocarbons from the catalyst. The spent catalyst containing coke leaves the stripping zone and passes into a regeneration zone where, in the presence of fresh regeneration gas, combustion of coke produces regenerated catalyst and flue gas containing carbon monoxide, carbon dioxide, water, nitrogen and perhaps a small quantity of oxygen. Usually the fresh regeneration gas is air but it could be air either enriched or deficient in oxygen. The amount of fresh regeneration gas to the regeneration zone is typically controlled by a predetermined temperature differential between the flue gas outlet section and either the dense bed temperature or the dilute phase temperature. Such a control scheme minimizes excess oxygen and allows only a small amount of afterburning, that is, only that amount characterized by the temperature differential, to take place. This control scheme does not permit controllable residual coke levels; indeed its purpose is to prevent getting "behind in burning" (of coke) by controlling to allow only a small amount of CO afterburning. When such a control scheme is used the amount of residual coke left on regenerated catalyst is largely a function of regeneration zone design, that is how well gas and solids are mixed, the number of stages used, the residence time and the resulting temperature. Typically regenerated catalyst will contain less than about 0.5 wt % coke and usually from about 0.15 to 0.35 wt % coke while spent catalyst entering the regeneration zone generally contains from about 0.5 to 1.5 wt % coke. Flue gas is separated from entrained regenerated catalyst by cyclone separation means located within the regeneration zone and separated flue gas is passed from

the regeneration zone typically to a carbon monoxide boiler where the chemical heat of carbon monoxide is recovered by combustion as a fuel for the production of steam. Regenerated catalyst which was separated from the flue gas is returned to the lower portion of the regeneration zone which is maintained as a dense bed of spent catalyst. Regenerated catalyst leaves this dense bed and as previously mentioned contacts the feedstock in a reaction zone. Generally, regenerated catalyst is not stripped of entrained flue gas prior to contacting the feed.

In a typical regeneration zone the spent catalyst is maintained in the bottom portion of the zone in one or more dense beds by limiting the superficial velocity of the incoming fresh regeneration gas. The superficial velocity is limited to the transport velocity, that is, the velocity above which large amounts of catalyst would be carried out of the dense bed to the cyclones. Typical velocities are therefore less than about 3 feed per second with 1.5 to 2.5 being the usual range.

Most of the total catalyst inventory of a fluid catalytic cracking process is contained in the regeneration zone and in view of the present industry trend toward short contact time reaction zones an even larger percentage of the total catalyst inventory is contained in the regeneration zone. The determination of the inventory in a typical regeneration zone is based upon the feed rate to the FCC process (or more specifically to the coke yield from that feed rate) and the superficial velocity. This coke yield anticipated from a desired feed rate determines the rate of the fresh regeneration gas to the regeneration zone. This gas rate at a limiting superficial velocity then determines the cross-sectional area of the regeneration zone. With a known catalyst density and height of the dense bed the inventory of the regeneration zone, and for practical purposes for the FCC process, is fixed. Catalyst residence times which result are generally from about 2 to 5 minutes with about 2 to 3 being the general range.

With the above description as a reference point, the concepts and advantages of our invention will be made more apparent.

In the process of our invention spent catalyst is regenerated to a controllable residual coke level and CO thereby produced is essentially completely burned to CO₂. Spent catalyst is passed into a first dense bed with a fresh regeneration gas stream controlled such as to produce regenerated catalyst having a desired residual coke level and partially spent regeneration gas. Regeneration gas and regenerated catalyst are transported from the first dense bed to a dilute phase transport riser where CO oxidation takes place and where the heat of combustion of the CO is transferred to the catalyst. Catalyst residence time in the transport riser is sufficiently short to preclude further substantial oxidation of the residual coke and additional production of CO. Catalyst and spent regeneration gas leaving the transport riser are separated and the regenerated catalyst is passed to a second dense bed where the catalyst can be stripped of entrained regeneration gas.

It is well known in the art that the level of residual coke on the regenerated catalyst has a great influence on the conversion and product yield distribution obtained in the reaction zone, especially in the short contact time, dilute phase reaction zones. For instance, at constant reactor temperature an increase in residual coke level on regenerated catalyst will reduce conversion and gasoline yield and increase cycle oil (fuel oil)

yield. A decrease in residual coke will increase conversion at the expense of fuel oil yield and increase gasoline yield. By returning to the reaction zone regenerated catalyst having very low residual coke, it is possible to change the yield structure in the direction of increased gas yields (LPG) at the expense of gasoline yield. The process of our invention not only recognizes the effect of residual coke on yields but allows the refiner to directly control the residual coke and thereby to control product yield. Thus the refiner can switch his operation from one maximizing fuel oil production to one maximizing gasoline yields or to one where maximum LPG is desired by directly changing the residual coke level. An example of these possible operations is shown in Table I below. Operations at intermediate coke levels are of course possible as well.

can be carried from the first dense bed into the dilute phase transport riser. Velocities contemplated for the dilute phase transport riser will be in the range of about 10 to 25 feet per second.

Since velocity is not only not limited to the transport velocity but is in fact 2 to 3 times the critical velocity, dramatic catalyst inventory reductions can now be achieved. As previously described, regeneration zone catalyst inventories are directly related to superficial velocities employed within the regeneration zone. Catalyst inventories using the process of this invention will be about 40 to 60 percent of those of present single or multistage regeneration processes. A moderately sized FCC process of the type presently in industry use will contain about 150 tons of catalyst. By using the regeneration process of this invention in an FCC process of

TABLE I

PRODUCT YIELDS AT DIFFERENT RESIDUAL COKE LEVELS

Type of FCC Operation	LPG	Gasoline	LCO (fuel oil)	Effect of residual coke at LPG conditions
Reactor temperature, °F	970	970	970	970
Feed Preheat Temp., °F	375	600	600	375
Regen. Cat. Delivery Temp., °F	1350	1350	1350	1350
Residual Coke on Regen. Catalyst, wt %	<0.02	0.11	0.62	0.62
Yield Distribution, % on Fresh Feed				
Conversion, LV%	85.7	75.0	64.4	77.0
C _x -C ₄ (LPG), LV%	38.2	26.3	20.4	31.4
Gasoline, LV% (380°F at 90% Over)	52.0	62.3	48.3	58.3
Light Cycle Oil, LV%	11.4	20.0	30.6	18.0
Clarified Slurry Oil, LV%	5.0	5.0	5.0	5.0
Coke, Wt %	6.8	3.8	5.3	6.8

The first operation shown in Table No. I is an LPG operation in which maximum LPG yield is produced at the indicated operating conditions using regenerated catalyst having very low residual coke. At the very low residual coke level of <0.02 wt % the high-activity regenerated catalyst returned to the reaction zone results in higher conversion and cracks part of the gasoline present to product this higher LPG yield. At an increased residual coke level of 0.11 wt % as shown in the gasoline operation, with the same reactor temperature, less active catalyst is returned to the reaction zone and as a result the conversion and LPG yield are reduced but the yield of gasoline is increased. As the residual coke level is further increased as shown in the LCO operation to about 0.6 wt % producing even less active catalyst, the conversion and yields of LPG and gasoline are decreased but the yield of light cycle oil (fuel oil) is increased. The effect of residual coke on product yields is further illustrated in the last column of Table No. I. Here the operating conditions are the same as those of the LPG operation except for the high residual coke level which is the same as that of the LCO operation. As shown, the conversion and LPG yield are less than those for the LPG operation while the gasoline and LCO yields are higher. This is attributable to less active catalyst caused by the higher residual coke level.

Since it is not intended in the process of our invention that catalyst in the first dense bed remain in that bed the superficial velocity of the fresh regeneration gas into that bed is not limited to the critical velocity. In the first dense bed superficial velocities will be in the range of about 3 to 10 feet per second so that catalyst

the same size, a refiner could save the initial investment represented by at least 75 tons of catalyst.

Catalyst makeup rates required to make up losses and maintain activity will also be reduced because such rates tend to be a percentage of the total catalyst inventory. Conversely, as previously mentioned, a heavier, more contaminated feed could be charged to an FCC process employing this invention without requiring a makeup rate in excess of that presently accepted on an FCC process using conventional regeneration techniques. Feedstocks would no longer have to be limited to relatively clean vacuum gas oils containing limited quantities of Conradson carbons, metals and nitrogen compounds. Higher molecular weight feed-stocks containing higher amounts of these contaminants could better be tolerated with no additional economic penalty. Because such feed streams require less processing, the refiner will realize an additional savings.

Better solid-gas contact and reduced residence times of both catalyst and regeneration gas are other advantages of the higher velocity. Higher velocities will produce more turbulent flow with better mixing and hence will result in more efficient regeneration. Because of this better gas-solid contact, along with the higher oxygen partial pressure and higher temperature, the rate of coke combustion will be increased and catalyst residence times can therefore be reduced. Catalyst residence times can be reduced from the present 2 to 5 minutes to less than 2 minutes and regeneration gas residence times can be reduced from about 20 seconds to less than 10 seconds. Better oxygen utilization will reduce the amount of fresh regeneration gas required

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which would therefore allow the regeneration zone pressure to be reduced. With shorter catalyst exposure time to high temperatures, catalyst activity will be prolonged and reduced fresh catalyst makeup rates can result.

From the standpoint of savings to be effected another important result of shorter catalyst residence time is that it now allows the steam stripping of flue gas components from the regenerated catalyst. It is present industry practice to strip only spent catalyst; spent catalyst from a reaction zone is stripped of absorbed and interstitial hydrocarbons before the catalyst is sent to the regeneration zone for the purpose of recovering valuable gasoline and light hydrocarbons that would otherwise be burned in the regeneration zone. Although it is known that prolonged exposure to steam will deactivate catalyst, steam is generally the preferred stripping medium rather than an inert gas because it can be condensed and separated before reaching the gas concentration section of the FCC process. The amounts used, however, are small and the short catalyst residence time, the low temperature and the coke on the catalyst minimize any deactivation. In spite of the fact that flue gas components are entrained by regenerated catalyst into the reaction zone and hence become part of a product stream, steam stripping of regenerated catalyst has generally not been practiced because of the longer catalyst residence time in the regeneration zone and the large catalyst inventory which is generally contained in a single dense bed. Exposure of this quantity of catalyst to steam for this longer period of time would increase the catalyst deactivation rate. Although an inert gas such as nitrogen would not tend to deactivate the catalyst, it would merely replace flue gas components, which are in themselves inert, and be carried into the product recovery and separation section of the FCC process. The following example and more detailed description will make clear this advantage of the process of this invention.

The effluent from the reaction zone of an FCC process actually contains not only hydrocarbons but steam from spent catalyst stripping and flue gas components in amounts from about 1 to 3 pounds per thousand pounds of catalyst circulated in the FCC process. In a typical FCC process of modest size, about 3,000,000 lb./hr. of catalyst will be circulated and will therefore entrain about 4500 lb./hr. of flue gas components into the reaction zone. This means that on a volume basis the reactor effluent will contain about 60,000 SCFH or 1,440,000 SCFD of flue gas components to be processed downstream of the reaction zone. The total reaction zone effluent is directed to the main column where it is separated into gas and unstabilized gasoline as an overhead product stream and various side cut product streams. This overhead product stream containing the flue gas components, light hydrocarbon gases, steam and gasoline is directed first to a main column overhead condenser where steam and gasoline are condensed and then to a main column overhead receiver where the gasoline and water are separated and the water is drawn off. In order to recover and separate light hydrocarbons and to stabilize the gasoline, the gas and unstabilized gasoline from the receiver are sent to a gas concentration process consisting primarily of a compressor, absorber columns, and fractionation columns along with associated equipment. To recover the light hydrocarbons the gas stream from the receiver, containing flue gas components, is first compressed to

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about 150 to 250 psig before being directed to the absorber columns. Light hydrocarbons, primarily C₃'s and C₄'s, are absorbed by heavier liquid hydrocarbon streams in the absorber columns leaving an unabsorbed lean gas containing the flue gas components as a product stream from the last absorber. A typical analysis of the off gas leaving the last absorber is as follows:

TABLE NO. II

	Absorber Off Gas
Composition (mol %)	
Carbon Dioxide	3.8
Oxygen + Argon	0.3
Nitrogen	22.8
Carbon Monoxide	4.5
Hydrogen	8.9
Methane	23.6
Ethylene	11.8
Ethane	12.5
Propylene	4.8
Propane	1.4
Isobutylene + 1-Butene	0.7
Cis-2-Butene	0.4
Trans-2-Butene	0.5
Isobutane	1.7
Normal Butane	0.5
Isopentane	0.7
Normal pentane	0.2
Total C ₄₊ and C ₅ Olefins	0.9
	100.0

As can be seen from the breakdown, the entrained flue gas components, nitrogen, carbon monoxide, and carbon dioxide, constitute about one-third of this product stream, (31.1%). Because of the short catalyst residence time employed in the process of our invention, stripping of the regenerated catalyst can be employed whereby most of this material would have been carried out of the regeneration zone rather than becoming a substantial part of this product stream. Without these components, substantial savings can be realized by using a less expensive gas plant, that is, one having smaller compressors and smaller absorbers along with other gas handling equipment. As well, the absorber off-gas which is generally used as a fuel would have a higher heat content.

When such stripping of regenerated catalyst in the second dense bed is provided, the bed volume will be so sized as to allow minimum catalyst exposure to steam. The bed volume will be sized to provide a catalyst residence time of less than about 1 minute and preferably less than about 30 seconds.

Temperatures employed in the first dense bed where coke oxidation takes place will be in the range of about 1250° to 1400° F and temperatures in the transport riser where combustion of CO and perhaps an additional combustible fluid takes place, will be in the range of about 1275° to about 1475° F. The exposure time of regenerated catalyst to these higher temperatures in the transport riser can be varied by the position of the fresh regeneration gas inlet into the transport riser. For shorter exposure times the inlet to the transport riser would be positioned nearer the riser outlet.

It is preferred that the temperature in the transport riser and the exposure time of the regenerated catalyst within the riser be such as to heat the catalyst which is passing through the riser to a temperature of at least 1250° F, preferably in the range of about 1250° F to 1400° F. Catalyst delivered to the reaction zone within this temperature range will not only promote rapid and efficient primary cracking of the feedstock but will reduce the feed preheat requirement. Normally feed to

the reaction zone is preheated to about 500°-700°F before contact with the regenerated catalyst to provide an additional heat input. Should the heat of combustion of CO within the transport riser be insufficient to heat the catalyst to the desired temperature, an additional combustible fluid may be added to and burned within the riser for this purpose. The combustible fluid could be fuel gas or any liquid hydrocarbon stream including the feedstock. This combustible fluid could as well be added to and burned in the first dense bed. The rate of fresh regeneration gas, controlled to oxidize coke to a predetermined residual coke level, would be increased slightly to take the combustible fluid into account.

A practical and economical advantage of essentially complete oxidation of CO within the regeneration zone is that it eliminates an air pollution problem without the necessity of a rather expensive CO boiler. Because of an increased worldwide awareness of the effects of a polluted environment on human health, particularly where an awareness is evidenced by more restrictive air pollution regulations, this advantage of our invention is very important. Since the heat of combustion of CO is transferred to the catalyst, an additional advantage is that the feed preheat requirement is reduced.

Pressures contemplated for use in the process of this invention are from about normal atmospheric pressure up to about 50 psig with the preferred range being 10 to 40 psig. Because of the improved gas-solid contact due to higher velocities, higher O₂ partial pressure due to more air for higher velocity, and the higher coke burning rate because of higher temperatures, lower pressures can be employed in the process of this invention with less penalty in coke burning capacity than for conventional designs. For this reason the process may reverse the present industry trend toward higher pressure regeneration processes. Less expensive vessels and air blower or compressors having lower pressure ratings therefore can be used.

Spent catalyst as used in this specification means catalyst withdrawn from a reaction zone because of reduced activity caused by coke deposits. Spent catalyst passing into the first dense bed can contain anywhere from a few tenths up to about 5 wt % of coke but typically in FCC operations spent catalyst removed from the reaction zone will contain from about 0.5 to about 1.5 wt % coke. The regenerated catalyst as used in this specification is catalyst from which essentially no more coke will be removed. For purposes of this specification catalyst leaving the first dense bed will be considered regenerated catalyst. The residence time of this catalyst within the dilute phase transport riser will be sufficiently short to preclude further coke oxidation. The regenerated catalyst will typically contain from about 0.5 wt % coke to as little as about 0.02 wt % coke depending upon the conversion level and yield structure desired from the reaction zone.

The term regeneration gas as used in the claims and in this specification shall mean, in a generic sense, any gas which is to contact catalyst or which has contacted catalyst within the regeneration zone. Specifically the term fresh regeneration gas shall include oxygen-containing gases such as air or oxygen enriched or deficient air which pass into the first dense bed of the regeneration zone to allow oxidation of coke on the spent catalyst therein. Partially spent regeneration gas shall refer to regeneration gas which has contacted catalyst within the first dense bed and which contains a reduced quantity of free oxygen. Typically the partially spent regen-

eration gas contains water, nitrogen, oxygen, carbon monoxide and carbon dioxide. Spent regeneration gas shall mean regeneration gas which contains substantially no carbon monoxide, from about a few tenths up to as much as 15 mole % free oxygen, carbon dioxide, nitrogen and water.

One or more gas-solids separation means may be utilized to separate spent regeneration gas from entrained regenerated catalyst. It is preferred, but not necessary to the process of this invention, that the catalyst separation means be connected to the dilute phase transport riser to receive and separate regeneration gas from entrained catalyst. Preferred separation means will be cyclone separators whose design and construction is well known to the art. A single cyclone may be used but preferably more than one of these cyclones will be used in parallel or in series flow arrangements to effect the desired degree of separation.

EXAMPLE

The following example is presented illustrating and comparing the process of our invention for the regeneration of spent catalyst with a conventional single stage regeneration method. In both cases a commercially available molecular sieve catalyst, which had previously been contacted with a vacuum gas oil under cracking conditions in a reaction zone and which had been subsequently steam stripped in a stripping zone, was regenerated with air. The coke contained 10.1 wt % hydrogen and coke on spent catalyst was about 0.9 wt %. The conventional regeneration zone contained a single dense bed in the bottom portion of the zone with a large dilute phase disengaging space positioned above the dense bed and the flue gas from this regeneration zone was burned in an external CO boiler. Analysis of the flue gas shown in Table No. III was for a sample removed from the flue gas line before the CO boiler. In the process of our invention CO (without an additional combustible fluid) was essentially completely combusted within the dilute phase transport riser of the regeneration zone. The flue gas sample for which the analysis is shown in Table No. III was taken just before flue gas exiting the regeneration zone was vented to the atmosphere.

TABLE NO. III

	REGENERATION PROCESS COMPARISON	
	Conventional	Present Invention
Temperatures, °F		
dense bed	1190	1250
dilute phase	1185	—
transport riser	—	1365
flue gas	1245	1360
second dense bed	—	1340
Pressure, psig	24.6	20
Net dry air to regenerator		
lb/hr	232,170	231,290
Dry air/coke, lb/lb	11.13	14.51
Coke yield, wt % of fresh feed (at 75% reaction zone conversion)	6.61	5.05
Coke on regenerated catalyst		
wt %	0.2	0.02
Controllable Coke	no	yes
Vessel size, diameter, ft.	25	16
Catalyst residence time, min	3	0.9
Gas residence time, sec	15.5	5.5
Superficial velocity, ft/sec		
dense bed	2.4	5
dilute phase	2.4	20
Flue Gas Analysis, vol %		
CO ₂	9.2	14.9
Argon	1.1	1.0
N ₂	79.2	82.3
CO	10.2	0 *

TABLE NO. III-continued

REGENERATION PROCESS COMPARISON		
	Conventional	Present Invention
O ₂	0.2	1.8
CH ₄	0.1	0
Regenerated catalyst stripping	no	yes
Feed preheat, °F	412	277
Catalyst Inventory, Tons	60	35

* Actual reading was 270 ppm vol.

The comparison of the process of our invention with the conventional single stage regeneration process first of all shows the higher temperatures and higher oxygen concentration of our invention. Although the inlet oxygen concentration in both cases is the same, (that of air) the outlet concentration from the process of our invention is 1.8 vol. %, compared to 0.2 vol. % from the conventional regeneration process. The higher regenerated catalyst temperature of our invention has reduced the catalyst circulation rate and thereby reduced the coke yield from 6.61 wt % to 5.05 wt %. This higher temperature has also reduced the feed preheat requirement by 135° F and, although not shown, will promote more rapid and efficient primary cracking of the feedstock in the reaction zone than will the lower temperature. This lower coke yield and more efficient cracking will result in more liquid product yield from the reaction zone. Because of the faster coke burning rate, caused by the higher temperature and higher oxygen concentration, our process has produced in this instance regenerated catalyst having a lower residual coke level. The product yield structure from the reaction zone, although not shown, will therefore be in the direction of lighter, more valuable products.

It must be noted also that in addition to making less coke yield at a particular conversion level, our process, in contrast to the conventional process, has the capability of controlling residual coke directly. It therefore offers the refiner a quick and convenient means of controlling the reaction zone conversion level and the production yield structure.

Table No. III flue gas analyses show that by combust-
ing CO within the regeneration zone our process has
essentially eliminated an air pollution problem without
requiring a CO boiler. It is to be noted also that this es-
sentially complete CO combustion, and the coke oxida-
tion to a lower residual coke level, was achieved using
less air (lb/hr) at a lower pressure than that for the con-
ventional process.

Additionally the catalyst inventory is about 40% less than that of the conventional process and therefore the size of the regeneration vessel is small. Although not shown, it is anticipated that because of the smaller inventory and shorter catalyst residence time, catalyst makeup rates will be less for our new process.

As indicated in Table III the regenerated catalyst is stripped in the present process but not in the conventional process. Although not shown in the table, the advantage of this stripping would be very apparent in a comparison of equipment sizes required in the gas concentration section.

The above specific example has been presented to illustrate preferred embodiments and advantages of the process of our invention. We do not intend to unreasonably limit the claims of our process by specific reference to sizes, flow rates, pressures, or analyses.

We claim as our invention:

1. A process for the regeneration of a zeolite-containing or an amorphous coke-contaminated spent hydrocarbon cracking catalyst, removed from a reaction zone, and the conversion of carbon monoxide, resulting from the oxidation of coke from said catalyst, to carbon dioxide, wherein the residual coke on the regenerated catalyst is directly controllable which process is operated within a pressure range from about normal atmospheric to about 50 psig, and comprises the steps of:
 - a. passing said catalyst and a fresh free oxygen-containing regeneration gas in a controlled amount to a first dense bed of fluidized particulate catalyst in a regeneration zone and oxidizing coke in said first dense bed maintained at oxidizing conditions including a temperature within the range of about 1250° F. to about 1400° F., a superficial gas velocity within the range of about 3 to about 10 feet per second, and a catalyst residence time of less than two minutes;
 - b. controlling the amount of fresh free oxygen-containing regeneration gas of step (a) passed into said dense bed to oxidize said coke to a predetermined residual coke level, to produce regenerated catalyst having a desired residual coke level and partially spent regeneration gas containing CO;
 - c. passing said regenerated catalyst and partially spent regeneration gas to a dilute phase transport riser communicating with and superimposed over said first dense bed and therein essentially completely converting CO to CO₂ at conversion conditions including a temperature within the range of about 1275° F. to 1475° F., a superficial gas velocity within the range of about 10 to 25 feet per second, and a residence time which precludes further substantial oxidation of residual coke and additional production of CO, to produce spent regeneration gas;
 - d. separating the thus-regenerated catalyst from said spent regeneration gas; and,
 - e. introducing said regenerated catalyst to a second dense bed of particulate material, from which said regenerated catalyst is returned to said reaction zone.
2. The process of claim 1 further characterized in that said regenerated catalyst is stripped of spent regeneration gas in said second dense bed.
3. The process of claim 1 further characterized in that the total residence time of regeneration gas in said first dense bed and dilute phase transport zone is less than 10 seconds.
4. The process of claim 1 further characterized in that said predetermined residual coke level is from about 0.02 wt. % to about 0.5 wt. % coke.
5. The process of claim 2 further characterized in that the catalyst residence time of said regenerated catalyst in said second dense bed is less than about 1 minute.

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